U.S. DEPARTMENT OF COMMERCE
National Technical Information Service

PB-264 270

Environmental Considerations of Selected Energy Conserving Manufacturing Process Options: Volume IV. Petroleum Refining Industry Report

Arthur D. Little, Inc, Cambridge, Mass.

Prepared for

Industrial Environmental Research Lab, Cincinnati, Ohio

Dec 76

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/7-76-034d	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE ENVIRONMENTAL CONSIDERATIONS OF SELECTED ENERGY CONSERVING MANUFACTURING PROCESS OPTIONS. Vol. IV, Petroleum Refining Industry Report.		5. REPORT DATE December 1976 (Issuing Date) 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S)		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Arthur D. Little, Inc. Acorn Park Cambridge, Massachusetts 02140		10. PROGRAM ELEMENT NO. 11. CONTRACT/GRANT NO. 68-03-2198
12. SPONSORING AGENCY NAME AND ADDRESS Industrial Environmental Research Laboratory Office of Research and Development U.S. Environmental Protection Agency Cincinnati, Ohio 45268		13. TYPE OF REPORT AND PERIOD COVERED Final 14. SPONSORING AGENCY CODE EPA/600/12

15. SUPPLEMENTARY NOTES Vol. III, EPA-600/7-76-034c, and V-XV, EPA-600/7-76-034e through EPA-600/7-76-034o, refer to studies of other industries as noted below; Vol.I, EPA-600/7-76-034a is the Industry Summary Report and Vol. II, EPA-600/7-76-034b is the Industry Priority Report.

16. ABSTRACT

This study assesses the likelihood of new process technology and new practices being introduced by energy intensive industries and explores the environmental impact of such changes.

Specifically Vol. IV deals with the petroleum refining industry and examines five options: (1) direct combustion of asphalt in process heaters and boilers, (2) hydrocracking of vacuum bottoms, (3) flexicoking of vacuum bottoms, (4) internal electrical power generation, (5) hydrogen generation by partial oxidation, all in terms of process economics and environmental/energy consequences. Vol. III and Vol. V-XV deal with the following industries: iron and steel, pulp and paper, olefins, ammonia, aluminum, textiles, glass, cement, chlor-alkali, phosphorus and phosphoric acid, copper, and fertilizers. Vol. I presents the overall summation and identification of research needs and areas of highest overall priority. Vol. II, prepared early in the study presents the methodology used to select industries.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Energy; Pollution; Industrial Wastes; Petroleum Refining	Manufacturing Processes Energy Conservation Asphalt H-oil	13B

DISTRIBUTION STATEMENT lease to public	19. SECURITY CLASS *(This Report)* unclassified	21. NO. OF PAGES 193
	20. SECURITY CLASS *(This page)* unclassified	22. PRICE

orm 2220-1 (9-73)

1a

PB 264 270

EPA

U.S. Environmental Protection Agency
Office of Research and Development

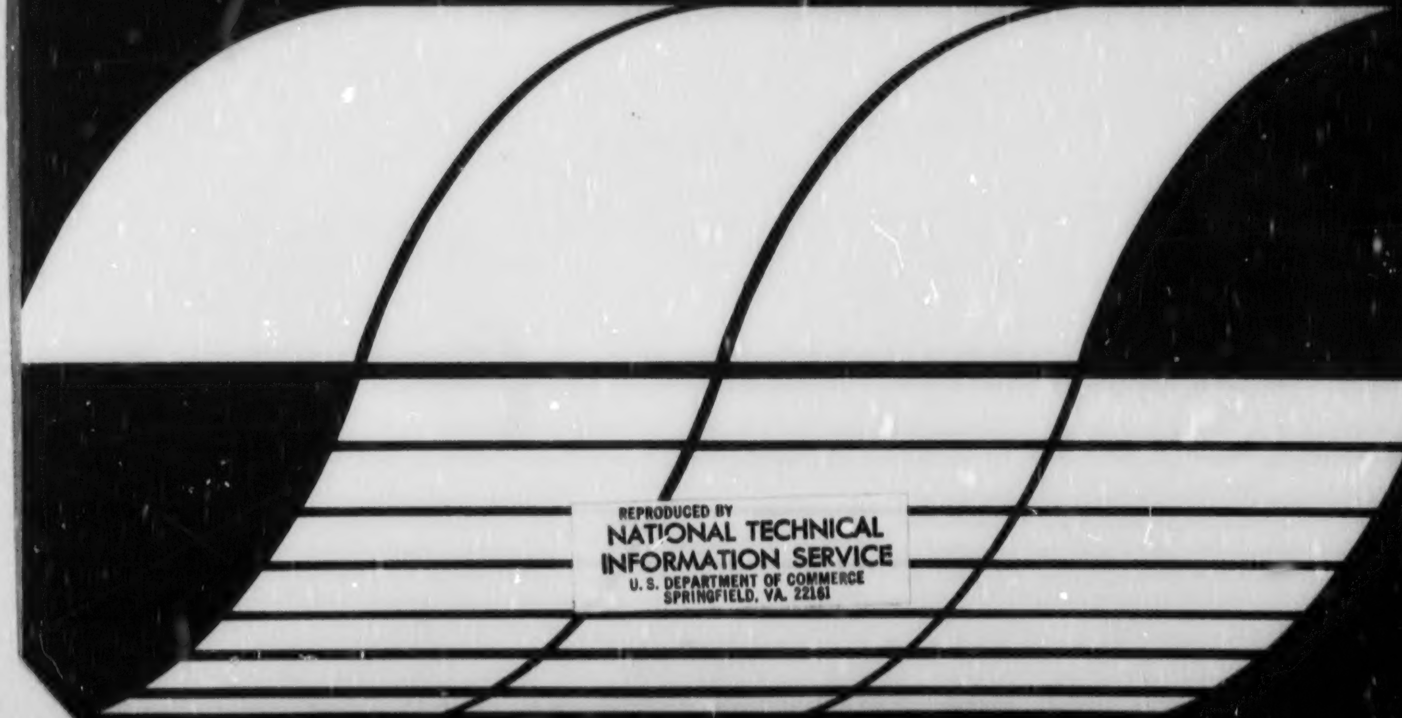
Industrial Environmental Research
Laboratory
Cincinnati, Ohio 45268

EPA-600/7-76-034d
December 1976

ENVIRONMENTAL CONSIDERATIONS OF SELECTED ENERGY CONSERVING MANUFACTURING PROCESS OPTIONS:

Vol. IV. Petroleum Refining Industry Report

Interagency
Energy-Environment
Research and Development
Program Report



REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into seven series. These seven broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and a maximum interface in related fields. The seven series are:

1. Environmental Health Effects Research
2. Environmental Protection Technology
3. Ecological Research
4. Environmental Monitoring
5. Socioeconomic Environmental Studies
6. Scientific and Technical Assessment Reports (STAR)
7. Interagency Energy-Environment Research and Development

This report has been assigned to the INTERAGENCY ENERGY-ENVIRONMENT RESEARCH AND DEVELOPMENT series. Reports in this series result from the effort funded under the 17-agency Federal Energy/Environment Research and Development Program. These studies relate to EPA's mission to protect the public health and welfare from adverse effects of pollutants associated with energy systems. The goal of the Program is to assure the rapid development of domestic energy supplies in an environmentally--compatible manner by providing the necessary environmental data and control technology. Investigations include analyses of the transport of energy-related pollutants and their health and ecological effects; assessments of, and development of, control technologies for energy systems; and integrated assessments of a wide range of energy-related environmental issues.

This document is available to the public through the National Technical Information Service, Springfield, Virginia 22161.

EPA-600/7-76-034d
December 1976

ENVIRONMENTAL CONSIDERATIONS OF SELECTED ENERGY CONSERVING MANUFACTURING PROCESS OPTIONS

Volume IV

PETROLEUM REFINING INDUSTRY REPORT

EPA Contract No. 68-03-2198

Project Officer

Herbert S. Skovronek
Industrial Pollution Control Division
Industrial Environmental Research Laboratory - Cincinnati
Edison, New Jersey 08817

INDUSTRIAL ENVIRONMENTAL RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

DISCLAIMER

This report has been reviewed by the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the U.S. Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This study, consisting of 15 reports, identifies promising industrial processes and practices in 13 energy-intensive industries which, if implemented over the coming 10 to 15 years, could result in more effective utilization of energy resources. The study was carried out to assess the potential environmental/energy impacts of such changes and the adequacy of existing control technology in order to identify potential conflicts with environmental regulations and to alert the Agency to areas where its activities and policies could influence the future choice of alternatives. The results will be used by the EPA's Office of Research and Development to define those areas where existing pollution control technology suffices, where current and anticipated programs adequately address the areas identified by the contractor, and where selected program reorientation seems necessary. Specific data will also be of considerable value to individual researchers as industry background and in decision-making concerning project selection and direction. The Power Technology and Conservation Branch of the Energy Systems-Environmental Control Division should be contacted for additional information on the program.

David G. Stephan
Director
Industrial Environmental Research Laboratory
Cincinnati

EXECUTIVE SUMMARY

Industry in the United States purchases about 27 quads* annually; approximately 40% of U.S. energy usage.** This energy is used for chemical conversion processing, steam raising, drying, space and process heating and cooling and miscellaneous other uses. In many industrial sectors, significant reductions in energy consumption can be achieved by either better "housekeeping" practices or the utilization of improved processes which are more efficient or result in upgrading of energy form values. This study focuses primarily on the latter category of energy conserving options.

A ranking system which considered both quantitative and qualitative factors, with emphasis on the total amount of energy utilized by an industry and the potential for energy conservation by a process change, was applied to various industries. Using this ranking method, the petroleum refining industry appeared in seventh place among the 13 industry segments being studied. The refining industry consumed about 2.96 quads in 1971 (see Table ES-1), 11% of the 27 quads purchased by all industries.

Petroleum refining is the manufacturing step in the production of hydrocarbon-based fuels. A large, modern refinery will produce a wide range of fuel products and contain a host of integrated processes which separate and convert crude petroleum into final products. As a result there are many potential opportunities for instituting energy conserving process changes within the overall crude oil refining process. A roster of potential energy conserving changes is provided in Appendix E. Through interviews with refining industry representatives, five alternatives likely to be implemented within the time frame of interest were selected for detailed evaluation.

- A-1 - Direct combustion of asphalt in process heaters and boilers
- A-2 - Hydrocracking of vacuum bottoms (H-Oil)
- A-3 - Flexicoking of vacuum bottoms
- B - Internal electrical power generation
- C - Hydrogen generation by partial oxidation

Each of the above options achieves an energy form value improvement as contrasted with an improvement in thermal efficiency.

---

*1 quad = 10^{15} Btu

**Purchased electricity value at an approximate fossil fuel equivalence of 10,500 Btu/kWh.

TABLE ES-1

SUMMARY OF 1971 ENERGY PURCHASED IN SELECTED INDUSTRY SECTOR

	Industry Sector	10^{15} Btu/Yr	SIC Code In Which Industry Found
1.	Blast furnaces and steel mills	3.49(1)	3312
2.	Petroleum refining	2.96(2)	2911
3.	Paper and allied products	1.59	26
4.	Olefins	0.984(3)	2818
5.	Ammonia	0.63(4)	287
6.	Aluminum	0.59	3334
7.	Textiles	0.54	22
8.	Cement	0.52	3241
9.	Glass	0.31	3211, 3221, 3229
10.	Alkalies and chlorine	0.24	2812
11.	Phosphorus and phosphoric acid production	0.12(5)	2819
12.	Primary copper	0.081	3331
13.	Fertilizers (excluding ammonia)	0.078	287

(1) Estimate for 1967 reported by FEA Project Independence Blueprint, p. 6-2, USGPO, November 1974.

(2) Includes captive consumption of energy from process byproducts (FEA Project Independence Blueprint)

(3) Olefins only, includes energy of feedstocks: ADL estimates

(4) Ammonia feedstock energy included: ADL estimates

(5) ADL estimates

Source: 1972 Census of Manufactures, FEA Project Independence Blueprint, USGPO, November 1974, and ADL estimates.

These process changes were evaluated within the context of "typical" existing refineries which were selected to take into account geographical differences in crude runs and product demands. The comparisons were made within the 1985 refinery environment which assumes complete phase out of natural gas from the refinery fuel system and complete conversion to the production of unleaded gasoline. (See Appendix C.)

The detailed assessment of these five alternatives resulted in the following energy related conclusions.

- The alternatives are generally net consumers of energy in a strictly thermodynamic sense. This is especially true when the energy associated with pollution control is included.
- Except for option A-3, the reduction in refinery efficiency is slight. For option A-3 refinery efficiency is reduced by two percentage points.
- The principal conservation benefit of these options is in terms of form value upgrading. This is accomplished by utilization or conversion of refinery residue streams which results in a net increase in the availability of higher valued fuels such as refinery gas and distillate range products. Table ES-2 summarizes the net product changes for each option.

The cost impact of pollution control associated with these process changes is summarized below.

- The implementation of the above process changes will have a small impact on the control of refinery wastewater. The alternatives evaluated do not change the characteristics of the treated effluent, or the type of treatment required to conform with regulatory requirements. In all cases, the wastewater loads represent increments to the treatment system, therefore the cost of treatment would not preclude the selection of any of the alternatives. At most, incremental wastewater treatment costs are 9% of the total operating cost (excluding feedstock and product values) for that option.
- The removal of sulfur to limits set for SO_x emissions has a significant cost impact on those options which involve the direct combustion of asphalt, i.e., options A-1 and B. In the case of option A-1, the operating cost for flue gas desulfurization (FGD) accounts for more than half of the total operating cost as shown in Table ES-3.
- The impact of EPA regulations on the residuum upgrading processes (H-Oil, Flexicoking, and Partial Oxidation) evaluated is not unreasonable in most cases. The additional control required represents an incremental change to the existing sulfur recovery system (see Table ES-3).

TABLE ES-2

CHANGES TO PRODUCT SLATES RESULTING FROM PROCESS CHANGES (BPD)

OPTIONS:	A-1 Direct Combustion		A-2 H-Oil		A-3 Flexicoking		B Power Generation		C Asphalt POX	
	Reduction	Increase	Reduction	Increase	Reduction	Increase	Reduction	Increase	Reduction	Increase
Products										
Asphalt (vac. btms.)	12,330	---	9,240	---	18,700	---	1,460	---	3,625	---
Tar	---	---	---	1,480	---	---	---	---	---	---
Residual oil (atm. btms.)	---		2,070	---	---	---	---	---	---	---
Heavy gas oil	---	9,640	---	3,820	---	8,746	---	---	---	---
Light gas oil	---		---	3,060	---	---	---	---	---	---
Naphtha	---	---	---	2,270	---	2,111	---	---	---	---
LPG (FOE)	---	---	---	540	---	350	---	---	---	---
Refinery gas (FOE)	---	2,690	800	---	---	4,489[1]	---	---	---	3,750
Coke	---	---	---	---	---	374	---	---	---	---
Electricity	---	---	---	---	---	---	---	1,460[2]	---	---
Total	12,330	12,330	12,110	11,170	18,700	16,070	1,460	1,460	3,625	3,750

---

[1] Includes displaced refinery gas and coker product gas.

[2] 1,460 ($6.3 \times 10^6 / 0.0105 \times 10^6$) = 876,000 kWh/day = 36.5 MW of capacity.

TABLE ES-3

SUMMARY OF ANNUAL POLLUTION CONTROL COSTS
($$10^3$)

Option	A-1	A-2	A-3	B	C
Water Pollution ($000/yr)					
With option change	5032	4613	5735	4963	4647
Base line	(5032)	(4541)	(5032)	(4919)	(4541)
Differential	0	74	703	44	106
Air Pollution ($000/yr)					
With option change	12,048[1]	2097	2840	1145	1145
Base line	9,997[2]	(1147)	(1130)	-	(1145)
Differential	2,051	950	1710	1145	0
Solid Waste ($000/yr)					
Differential	-	2	230	-	4
Total differential ($000/yr)	2051	1026	2643	1189	110
Cost of pollution control, $000/day	6.2	3.1	8.0	3.6	0.3
Operating costs for options, $000/day	5.6	19.3	22.6	20.4	38.6
Total PC & operating cost, $000/day	11.8	22.4	30.6	24.0	38.9
Control cost, % of total	53	14	26	15	<1

---

[1] Assumes pollution control by FGD @ $\$0.47/10^6$ Btu.

[2] Assumes pollution control by HDS and blending @ $\$0.39/10^6$ Btu.

EPA policies regarding sulfur emissions from refineries could certainly have an effect upon how the refiners manage their internal energy requirements. The present approach is to blend low sulfur distillate oils with the heavy bottoms to meet an overall sulfur specification. In so doing, the yield of distillate oil product may be reduced. Option A-1 presents an alternative which would make more light fuel oil available to consumers, however, the FGD pollution control technology available is too costly for practial implementation. Hence, the impact of EPA regulations upon the management of the refinery fuel balance in this case has an affect on the choice of options.

The same problem is encountered with the integrated electrical/steam system (option B). First, it is the only option which might be undertaken to conserve total energy in the refinery, not just upgrade form value or higher product value. Hence, some movement in this direction may be mandated by FEA, regardless of the economics. Second with this motivation, it underlines the great need for small source FGD developments, perhaps utilizing typical refinery resources such as CO or H_2.

The choice of alternates for heavy resid conversion is particularly dependent upon the type of crudes processed at a given refinery and the particular markets for asphalt, coke and distillate fuels. Furthermore, the impact of pollution regulations is not particularly great for these options, consequently, pollution regulations are not seen as an impediment in the application of these technologies. However, less costly small source FGD technology would also benefit these form value upgrading schemes.

As a result of this assessment, two technology areas have been identified which through added research could increase energy conservation in terms of form value availability. By far the most important research area would have as an objective improving the reliability and reducing the cost of flue gas desulfurization, especially in the size range below 50 MW equivalent. Such a development should make maximum use of refinery resources, such as carbon monoxide, hydrogen, amine plants, or Claus units. The second is concerned with developing rugged hydrocracking catalysts which can withstand the poisons normally present in petroleum residues, or alternatively an economical residual demetalization process. The objective in this case would be to increase space velocity (more activity) and yields. Technology improvements in these areas would allow flexibility in refinery fuel management and increase the yields of clean fuels from refineries without sacrificing environmental quality.

This report was submitted in partial fulfillment of contract 68-03-2198 by Arthur D. Little, Inc. under sponsorship of the U.S. Environmental Protection Agency. This report covers a period from June 9, 1975 to February 9, 1976.

TABLE OF CONTENTS

Page

FOREWORD iii
EXECUTIVE SUMMARY iv
List of Figures xiii
List of Tables xiv
Acknowledgments xix
Conversion Table xxi

I. INTRODUCTION 1

A. BACKGROUND 1
B. CRITERIA FOR INDUSTRY SELECTION 1
C. CRITERIA FOR PROCESS SELECTION 2
D. SELECTION OF PETROLEUM REFINING INDUSTRY PROCESS OPTIONS 3

II. FINDINGS, CONCLUSIONS, AND RECOMMENDATIONS 6

A. CHANGES STUDIED 6
B. IMPACTS ON WASTEWATER CONTROL 6
C. ENERGY CONSERVATION POSSIBILITIES 7

III. PETROLEUM REFINING INDUSTRY OVERVIEW 9

A. INDUSTRY DESCRIPTION 9
B. ECONOMIC OUTLOOK 11

IV. COMPARISON OF CURRENT AND ALTERNATIVE PROCESSES 14

A. CRITERIA FOR ASSESSING OPTIONS 14

1. Probability of Implementation 14
2. Pollution Control 14
3. Energy Considerations 14
4. Feedstock Availability 15
5. Market Considerations 15
6. Process Economics 16

B. SELECTION OF STUDY ALTERNATIVES 16

1. Option A: Utilization of Asphalt to Increase Yields of Higher Value Products 16

Preceding page blank

TABLE OF CONTENTS (Cont.)

Page

2. Option B: Internal Power Generation Using Asphalt 18
3. Option C: Alternative Methods for Hydrogen Generation 19

C. REFINERY BASE LINE AND PROCESSING ALTERNATIVES 21

1. Refinery Base Line Used for Comparison 21
2. Process Change A-1: Direct Combustion of Asphalt in Process Heaters and Boilers 32
3. Process Change A-2: Hydrocracking of Heavy Bottoms 43
4. Process Change A-3: Flexicoking 52
5. Process Change B: On-Site Electric Power by Combustion of Vacuum Bottoms 67
6. Process Change C: High-Purity Hydrogen Production via Partial Oxidation of Asphalt 78

V. IMPACTS OF POTENTIAL INDUSTRY CHANGES 90

A. IMPACT OF PROCESSING OPTIONS 90

1. Process/Pollution Control/Energy Effects 90
2. Cost Impact 92

B. SYSTEMS IMPLICATIONS 92

1. Direct Combustion of Asphalt 96
2. On-Site Power Generation 96
3. Upgrading of Heavy Residues (H-Oil, Flexicoking, Asphalt POX) 96

C. FACTORS AFFECTING PROBABILITY OF CHANGE 96

1. Direct Combustion of Asphalt 96
2. Power Generation 97
3. Upgrading of Asphalt 97

D. AREAS FOR RESEARCH 97

APPENDIX A - INDUSTRY STRUCTURE 99

APPENDIX B - PRESENT TECHNOLOGY 111

APPENDIX C - BASE LINE ENERGY USE 113

APPENDIX D - CURRENT POLLUTION PROBLEMS AND EFFECTIVENESS OF AVAILABLE POLLUTION CONTROL TECHNOLOGY 134

APPENDIX E - ROSTER OF PROCESS OPTIONS OR CHANGES IN INDUSTRIAL PRACTICE 162

APPENDIX F - GLOSSARY 167

LIST OF FIGURES

Number		Page
III-1	Petroleum Administration for Defense (PAD) Districts	10
IV-1	Vacuum Residue Distribution System	34
IV-2	Pollution Source Identification - Asphalt Combustion	37
IV-3	H-Oil	44
IV-4	Identification of Pollution Sources from Hydrocracking	47
IV-5	Flow Diagram and Volume Flows for Flexicoking	56
IV-6	Block Flow Diagram for Flexicoking and Identification of Pollution Sources from Flexicoking	60
IV-7	Electric Power from Asphalt	69
IV-8	Pollution Source Identification: Electric Power from Asphalt Combustion	72
IV-9	Hydrogen via Partial Oxidation	79
IV-10	Pollution Source Identification: Partial Oxidation	83
A-1	Petroleum Administration for Defense (PAD) Districts	101
C-1	East Coast Cluster Calibration	116
C-2	Louisiana Gulf Coast Cluster Model Calibration	117
C-3	West Coast Cluster Model Calibration	118
D-1	Capital Cost for a Typical Sulfur - Recovery Technology	153
D-2	Operating Costs for Typical Sulfur - Recovery Technology	154
D-3	Capital and Operating Costs vs. Heat Input Rate for Scrubber System	156

LIST OF TABLES

Number		Page
I-1	Summary of 1971 Energy Purchased in Selected Industry Sectors	3
III-1	Survey of U.S. Refining Industry by Capacity and Number of Plants	9
III-2	Survey of Operating Refineries in the United States - 1962-1972	12
IV-1	Flexicoker Yields From Various Residua	19
IV-2	Anticipated Shifts in Refinery Energy Usage Caused by Natural Gas and TEL Phase-Out	23
IV-3	Major Airborne Emissions From Base Line Refinery (East Coast)	24
IV-4	Major Airborne Emissions From Base Line Refinery (Gulf Coast)	25
IV-5	Major Airborne Emissions From Base Line Refinery (West Coast)	26
IV-6	Base Case Refinery - Characteristics of Raw and Treated Process Wastewater	28
IV-7	Estimation of Base Case Solid Waste Generation	29
IV-8	Summary of Base Line Wastewater Treatment Costs	30
IV-9	Existing Refinery Capital Investments and Operating Costs	33
IV-10	Comparison of Refinery Energy Balance With and Without Direct Combustion of Asphalt	35
IV-11	Asphalt Combustion Pollution Profile	38
IV-12	Example of SO_2 Control for Process Heater or Boiler	40
IV-13	Production Costs: Option A-1 (Direct Combustion)	42
IV-14	Comparison of Refinery Energy Balance With and Without Heavy-ends Hydrocracking	46

LIST OF TABLES (Cont.)

Number		Page
IV-15	Summary of Addition to Base Line Emissions Due to Heavy-ends Hydrocracking	48
IV-16	West Coast Refinery With Heavy-bottoms Hydrocracking Process Wastewater Treatment Costs	50
IV-17	Sulfur Control Costs for Base Line and With Hydrocracking (Option A-2)	51
IV-18	Production Costs: Option A-2 (Resid Hydrocracking)	53
IV-19	Products From Heavy-ends Hydrocracking	54
IV-20	Comparison of Product Slates For the West Coast Base Line With the Incorporation of Heavy-ends Hydrocracking (1000 BPD)	54
IV-21	Summary of Refinery Energy Balance With and Without Flexicoking	58
IV-22	Summary of Additions to Base Line Emissions Due to Flexicoking	59
IV-23	Water Analysis Flexicoking Sour Water	61
IV-24	East Coast Refiner With Flexicoking Process Wastewater Treatment Costs	62
IV-25	East Coast Refinery With Flexicoking Cooling Tower Blowdown Wastewater Treatment Costs	63
IV-26	Sulfur Control Costs for Flexicoking (Option A-3)	65
IV-27	Production Costs: Option A-3 (Flexicoking)	66
IV-28	Comparison of Product Slates With and Without Flexicoking	68
IV-29	Comparison of Refinery Energy Balance With and Without On-site Power Generation	70
IV-30	Electric Power From Asphalt - Pollution Profile	73
IV-31	Gulf Coast Refinery With On-site Power Generation Cooling Tower Blowdown Wastewater Treatment Costs	74
IV-32	Sulfur Control Costs of Electric Power Generation (Option B)	76
IV-33	Production Costs: Option B (Steam/Electric)	77
IV-34	Comparison of Refinery Energy Balance With and Without Hydrogen Production via Partial Oxidation of Asphalt	81

LIST OF TABLES (Cont.)

Number		Page
IV-35	Partial Oxidation for Hydrogen Production - Pollution Profile	84
IV-36	West Coast Refinery - Hydrogen Production by Partial Oxidation Process Wastewater Treatment Costs	85
IV-37	West Coast Refinery With Partial Oxidation Hydrogen Production Cooling Tower Blowdown Wastewater Treatment Costs	86
IV-38	Sulfur Control Costs for Hydrotreating (Option A-2) and Partial Oxidation (Option A-3)	87
IV-39	Production Costs: Option C (Partial Oxidation)	89
V-1	Changes to Product Slates Resulting From Process Changes	91
V-2	Change in Refinery Energy Usage Resulting From Process Changes	93
V-3	Energy Consumed by Air Pollution Control (APC) Systems	94
V-4	Summary of Annual Pollution Control Costs	95
A-1	Survey of Operating Refineries in the United States - 1962-1972	100
A-2	U.S. Petroleum Refining Capacity and Actual Crude Runs Average January 1 Capacity in Given Year and January 1 Capacity in the Following Year	102
A-3	Historical Refining Capacity Data - Total United States	104
A-4	New Refineries, Expansions and Reactivations Scheduled in the United States by PAD Districts	105
A-5	Refineries Planned but not Constructed due to Opposition on Environmental Grounds	108
A-6	Projection of U.S. Primary Energy Supplies with Oil as the Balancing Fuel	110
A-7	Forecast of U.S. Product Demand	110
C-1	Refineries Simulated by Cluster Models	114
C-2	Summary of East Coast Refinery Operation (1974)	119

LIST OF TABLES (Cont.)

Number		Page
C-3	Summary of Gulf Coast Refinery Operation	120
C-4	Summary of West Coast Refinery Operation	121
C-5	Summary of East Coast Refinery Energy Consumption (1974)	122
C-6	Summary of Gulf Coast Refinery Energy Consumption (1974)	123
C-7	Summary of West Coast Refinery Energy Consumption (1974)	124
C-8	Summary of East Coast Refinery Operation (1985)	126
C-9	Summary of Gulf Coast Refinery Operation (1985)	127
C-10	Summary of West Coast Refinery Operation (1985)	128
C-11	Summary of East Coast Refinery Energy Consumption (1985)	129
C-12	Summary of Gulf Coast Refinery Energy Consumption (1985)	130
C-13	Summary of West Coast Refinery Energy Consumption (1985)	131
C-14	Anticipated Shifts in Refinery Energy Usage Caused by Natural Gas and TEL Phase-Out	132
C-15	Capital Investments and Operating Costs	133
D-1	Typical Base Case Raw Refinery Waste Loads	136
D-2	Base Case Refinery - Characteristics of Raw and Treated Process Wastewater	138
D-3	Process Wastewater Treatment Costs: East Coast Base Case Refinery	141
D-4	Cooling Tower Blowdown Wastewater Treatment Costs: East Coast Base Case Refinery	142
D-5	Process Wastewater Treatment Costs: Gulf Coast Base Case Refinery	143
D-6	Cooling Tower Blowdown Wastewater Treatment Costs: Gulf Coast Base Case Refinery	144
D-7	Process Wastewater Treatment Costs: West Coast Base Case Refinery	145

LIST OF TABLES (Cont.)

Number		Page
D-8	Cooling Tower Blowdown Wastewater Treatment Cost: West Coast Refinery	146
D-9	Emission Factors in Petroleum Refineries	147
D-10	Anticipated Shifts in Refinery Energy Usage Caused by Natural Gas and TEL Phase-Out	149
D-11	Major Airborne Emissions from Base Line Refinery Location: East Coast Year: 1985	150
D-12	Major Airborne Emissions from Base Line Refinery Location: Gulf Coast Year: 1985	151
D-13	Air Pollution Control Cost of Large Combustion Sources	157
D-14	Petroleum Refinery Solid Wastes--Sources and Characteristics	160
D-15	Base Case Refinery - Characteristics of Raw and Treated Process Wastewater	161
F-1	Weight Conversions	169
F-2	Volume Conversions	169
F-3	Nomenclature	170

ACKNOWLEDGMENTS

This study could not have been accomplished without the support of a great number of people in government agencies, industry, trade associations and universities. Although it would be impossible to mention each individual by name, we would like to take this opportunity to acknowledge the particular support of a few such people.

Dr. Herbert S. Skovronek, Project Officer, was a valuable resource to us throughout the study. He not only supplied us with information on work presently being done in other branches of EPA and other government agencies, but served as an indefatigable guide and critic as the study progressed. His advisors within EPA, FEA, DOC, and NBS also provided us with insights and perspectives valuable for the shaping of the study.

During the course of the study we also had occasion to contact many individuals within industry and trade associations. Where appropriate we have made reference to these contacts within the various reports. Frequently, however, because of the study's emphasis on future developments with comparative assessments of new technology, information given to us was of a confidential nature or was supplied to us with the understanding that it was not to be credited. Therefore, we extend a general thanks to all those whose comments were valuable to us for their interest in and contribution to this study.

Finally, because of the broad range of industries covered in this study, we are indebted to many people within Arthur D. Little, Inc. for their participation. Responsible for the guidance and completion of the overall study were Mr. Henry E. Haley, Project Manager; Dr. Charles L. Kusik, Technical Director; Mr. James I. Stevens, Environmental Coordinator; and Ms. Anne B. Littlefield, Administrative Coordinator.

Members of the environmental team were Dr. Indrakumar L. Jashnani, Mr. Edmund H. Dohnert and Dr. Richard Stephens (consultant).

Within the individual industry studies we would like to acknowledge the contributions of the following people.

Iron and Steel:	Dr. Michel R. Mounier, Principal Investigator Dr. Krishna Parameswaran
Petroleum Refining:	Mr. R. Peter Stickles, Principal Investigator Mr. Edward Interess Mr. Stephen A. Reber Dr. James Kittrell (consultant) Dr. Leigh Short (consultant)

Pulp and Paper:	Mr. Fred D. Iannazzi, Principal Investigator Mr. Donald B. Sparrow Mr. Edward Myskowski (consultant) Mr. Karl P. Fagans Mr. G. E. Wong
Olefins:	Mr. Stanley E. Dale, Principal Investigator Mr. R. Peter Stickles Mr. J. Kevin O'Neill Mr. George B. Hegeman
Ammonia:	Mr. John L. Sherff, Principal Investigator Ms. Nancy J. Cunningham Mr. Harry W. Lambe
Aluminum:	Mr. Richard W. Hyde, Principal Investigator Ms. Anne B. Littlefield Dr. Charles L. Kusik Mr. Edward L. Pepper Mr. Edwin L. Field Mr. John W. Rafferty
Textiles:	Dr. Douglas Shooter, Principal Investigator Mr. Robert M. Green (consultant) Mr. Edward S. Shanley Dr. John Willard (consultant) Dr. Richard F. Heitmiller
Cement:	Dr. Paul A. Huska, Principal Investigator Ms. Anne B. Littlefield Mr. J. Kevin O'Neill
Glass:	Dr. D. William Lee, Principal Investigator Mr. Michael Rossetti Mr. R. Peter Stickles Mr. Edward Interess Dr. Ravindra M. Nadkarni
Chlor-Alkali:	Mr. Roger E. Shamel, Principal Investigator Mr. Harry W. Lambe Mr. Richard P. Schneider
Phosphorus/ Phosphoric Acid:	Mr. William V. Keary, Principal Investigator Mr. Harry W. Lambe Mr. George C. Sweeney Dr. Krishna Parameswaran
Primary Copper:	Dr. Ravindra M. Nadkarni, Principal Investigator Dr. Michel R. Mounier Dr. Krishna Parameswaran
Fertilizers:	Mr. John L. Sherff, Principal Investigator Mr. Roger Shamel Dr. Indrakumar L. Jashnani

ENGLISH-METRIC (SI) CONVERSION FACTORS

To Convert From	To	Multiply By
Acre	Metre^2	4,046
Atmosphere (normal)	Pascal	101,325
Barrel (42 gal)	Metre^3	0.1589
British Thermal Unit	Joule	1,055
Centipoise	Pascal-second	0.001
Degree Fahrenheit	Degree Celsius	$t_c^\circ = (t_F^\circ - 32)/1.8$
Degree Rankine	Degree Kelvin	$t_K^\circ = t_R^\circ/1.8$
Foot	Metre	0.3048
Foot^3/minute	Metre^3/sec	0.0004719
Foot^3	Metre^3	0.02831
Foot^2	Metre^2	0.09290
Foot/sec	Metre/sec	0.3048
Foot^2/hr	Metre^2/sec	0.00002580
Gallon (U.S. liquid)	Metre^3	0.003785
Horsepower (550 ft-lbf/sec)	Watt	745.7
Horsepower (electric)	Watt	746.0
Horsepower (metric)	Watt	735.5
Inch	Metre	0.02540
Kilowatt-hour	Joule	3.60×10^6
Litre	Metre^3	1.000×10^{-3}
Micron	Metre	1.000×10^{-6}
Mil	Metre	0.00002540
Mile (U.S. statute)	Metre	1,609
Poise	Pascal-second	0.1000
Pound force (avdp)	Newton	4.448
Pound mass (avdp)	Kilogram	0.4536
Ton (assay)	Kilogram	0.02916
Ton (long)	Kilogram	1,016
Ton (metric)	Kilogram	1,000
Ton (short)	Kilogram	907.1
Tonne	Kilogram	1,000

Source: American National Standards Institute, "Standard Metric Practice Guide," March 15, 1973. (ANS72101-1973) (ASTM Designation E380-72)

I. INTRODUCTION

A. BACKGROUND

Industry in the United States purchases about 27 quads* annually, approximately 40% of total national energy usage.** This energy is used for chemical processing, raising steam, drying, space cooling and heating, process stream heating, and miscellaneous other purposes.

In many industrial sectors energy consumption can be reduced significantly by better "housekeeping" (i.e., shutting off standby furnaces, better thermostat control, elimination of steam and heat leaks, etc.) and greater emphasis on optimization of energy usage. In addition, however, industry can be expected to introduce new industrial practices of processes either to conserve energy or to take advantage of a more readily available or less costly fuel. Such changes in industrial practices may result in changes in air, water or solid waste discharges. The EPA is interested in identifying the pollution loads of such new energy-conserving industrial practices or processes and in determining where additional research, development, or demonstration is needed to characterize and control the effluent streams.

B. CRITERIA FOR INDUSTRY SELECTION

In the first phase of this study we identified industry sectors that have a potential for change, emphasizing those changes which have an environmental/energy impact.

Industries were eliminated from further consideration within this assignment if the only changes that could be envisioned were:

- energy conservation as a result of better policing or "housekeeping,"
- better waste heat utilization,
- fuel switching in steam raising, or
- power generation.

After discussions with the EPA Project Officer and his advisors, industry sectors were selected for further consideration and ranked using:

---

*1 quad = 10^{15} Btu

**purchased electricity valued at an approximate fossil fuel equivalent of 10,500 Btu/kWh

- Quantitative criteria based on the gross amount of energy (fossil fuel and electric) purchased by industry sector as found in U.S. Census figures and from information provided from industry sources. The petroleum refining industry purchased 2.96 quads out of the 12.14 quads purchased in 1971 by the 13 industries selected for study, or 11% of the 27 quads purchased by all industry (see Table I-1).

- Qualitative criteria relating to probability and potential for process change, and the energy and effluent consequences of such changes.

In order to allow for as broad a coverage of technologies as possible, we then reviewed the ranking, eliminating some industries in which the process changes to be studied were similar to those in another industry planned for study. We believe the final ranking resulting from these considerations identifies those industry sectors which show the greatest possibility of energy conservation via process change. Further details on this selection process can be found in the Industry Priority Report prepared under this contract (Volume II). On the basis of this ranking method, the petroleum refining industry appeared in seventh place among the 13 industrial sectors listed.

C. CRITERIA FOR PROCESS SELECTION

In this study we have focused on identifying changes in the primary production processes which have clearly defined pollution consequences. In selecting those to be included in this study, we have considered the needs and limitations of the EPA as discussed more completely in the Industry Priority Report mentioned above. Specifically, energy conservation has been defined broadly to include, in addition to process changes, conservation of energy or energy form (gas, oil, coal) by a process or feedstock change. Natural gas has been considered as having the highest energy form value followed in descending order by oil, electric power, and coal. Thus, a switch from gas to electric power would be considered energy conservation because electric power could be generated from coal, existing in abundant reserves in the United States in comparison to natural gas. Moreover, pollution control methods resulting in energy conservation have been included within the scope of this study. Finally, emphasis has been placed on process changes with near-term rather than long-term potential within the 15-year span of time of this study.

In addition to excluding from consideration better waste heat utilization, "housekeeping," power generation, and fuel switching, as mentioned above, certain options have been excluded to avoid duplicating work being funded under other contracts and to focus this study more strictly on "process changes." Consequently, the following have also not been considered to be within the scope of work:

- Carbon monoxide boilers (however, unique process vent streams yielding recoverable energy could be mentioned);

- Fuel substitution in fired process heaters;

- Mining and milling, agriculture, and animal husbandry;

TABLE I-1

SUMMARY OF 1971 ENERGY PURCHASED IN SELECTED INDUSTRY SECTORS

	Industry Sector	10^{15} Btu/Yr	SIC Code In Which Industry Found
1.	Blast furnaces and steel mills	3.49(1)	3312
2.	Petroleum refining	2.96(2)	2911
3.	Paper and allied products	1.59	26
4.	Olefins	0.984(3)	2818
5.	Ammonia	0.63(4)	287
6.	Aluminum	0.59	3334
7.	Textiles	0.54	22
8.	Cement	0.52	3241
9.	Glass	0.31	3211, 3221, 3229
10.	Alkalies and chlorine	0.24	2812
11.	Phosphorus and phosphoric acid production	0.12(5)	2819
12.	Primary copper	0.081	3331
13.	Fertilizers (excluding ammonia)	0.078	287

(1) Estimate for 1967 reported by FEA Project Independence Blueprint, p. 6-2, USGPO, November 1974.

(2) Includes captive consumption of energy from process byproducts (FEA Project Independence Blueprint)

(3) Olefins only, includes energy of feedstocks: ADL estimates

(4) Amonia feedstock energy included: ADL estimates

(5) ADL estimates

Source: 1972 Census of Manufactures, FEA Project Independence Blueprint, USGPO, November 1974, and ADL estimates.

- Substitution of scrap (such as reclaimed textiles, iron, aluminum, glass, and paper) for virgin materials;
- Production of synthetic fuels from coal (low- and high-Btu gas, synthetic crude, synthetic fuel oil, etc.); and
- All aspects of industry-related transportation (such as transportation of raw material).

Within the refining industry, certain exceptions were considered since fuel substitution is integrally related to and can have a significant impact on the amount of product produced.

D. SELECTION OF PETROLEUM REFINING INDUSTRY PROCESS OPTIONS

Within each industry, the magnitude of energy use was an important criterion in judging where the most significant energy savings might be realized, since reduction in energy use reduces the amount of pollution generated in the energy production step. Guided by this consideration, candidate options for in-depth analysis were identified from the major energy consuming process steps with known or potential environmental problems.

After developing a list of candidate process options, we assessed subjectively

- pollution or environmental consequences of the process change,
- probability or potential for the change, and
- energy conservation consequences of the change.

Even though all of the candidate process options were large energy users, there was wide variation in energy use and estimated pollution loads between options at the top and bottom of the list. A modest process change in a major energy consuming process step could have more dramatic energy consequences than a more technically significant process change in a process step whose energy consumption is rather modest. For the lesser energy-using process steps process options were selected for in-depth analysis only if a high probability for process change and pollution consequences was perceived.

Because of the time and scope limitations for this study, we have not attempted to prepare a comprehensive list of process options or to consider all economic, technological, institutional, legal or other factors affecting implementation of these changes. Instead we have relied on our own background experience, industry contacts, and the guidance of the Project Officer and EPA advisors.

Reconciling such difficulties with the desire to cover as wide a spectrum as possible of the consequences of process change, the following candidates were considered within the petroleum refining industry:

- Burning asphalt in heaters and boilers.
- Generating hydrogen from asphalt by partial oxidation.
- Hydrocracking vacuum bottoms.
- Flexicoking vacuum bottoms.
- Generating electricity by burning asphalt.
- Blending methanol into gasoline.
- Computer optimization of hydrogen balance.
- Generating hydrogen from coal (Texaco POX).
- Upgrading byproduct hydrogen purity by membrane separation.
- FCC UltraCatalytic process (Amoco).

After discussion with the EPA Project Officer, his advisors, and industry representatives, ADL chose the first five from this list for analysis because:

- high crude oil prices and limited markets for heavy bottoms will lead to more conversion of asphalt into higher quality fuel products;
- there will be more emphasis on optimizing the allocation of fuels to meet refinery energy needs; and
- future electricity rate structures for industrial power may favor internal power generation.

The remaining options are only qualitatively discussed based on readily available information and discussions with industry representatives.

1. Method of Analysis

In undertaking this study, the petroleum refining industry description is based on 1974, the latest representative year for which we had good statistical information.

The impacts of the selected process changes were compared within the context of reference refineries with 1985 operating characteristics. A 1985 time frame was used to allow for certain fundamental changes in refining which have already begun and will be accomplished by then. These changes include the phase-out of natural gas as refinery fuel and the phase-out of TEL in the gasoline pool. Three refinery cluster models typifying East, Gulf, and West Coast refinery operation were utilized to assess the impact of processing changes in different regions.

Recognizing that capital costs and energy costs have escalated rapidly in the past few years and have greatly distorted the traditional basis for making cost comparison, we believe that the most meaningful economic assessment of new process technology can only be made by using 1975 cost data to the extent possible. Consequently, in estimating operating costs we have developed costs representative of the first half of 1975 using constant 1975 dollars for our comparative analysis of future process options.

II. FINDINGS, CONCLUSIONS, AND RECOMMENDATIONS

A. CHANGES STUDIED

The five industry/process changes assessed in this study are briefly:

A-1. Direct combustion of asphalt in heaters/boilers,

A-2. Asphalt conversion by hydrocracking (H-OIL),

A-3. Asphalt conversion by flexicoking,

B. Internal power generation, and

C. Hydrogen generation by partial oxidation.

We evaluated these options within the context of "typical" existing refineries representing actual refinery clusters located in the PAD* districts selected. Each of these options achieves an energy form value improvement.

B. IMPACTS ON WASTEWATER CONTROL

Within the framework of the comparisons set forth, we concluded that the implementation of these changes will have a small impact on the control of refinery wastewater. The effect on the characteristics of the treated effluent, the type of treatment required to conform with regulatory requirements, and the cost of wastewater treatment would not preclude the selection of any of the alternatives. Option A-3 has the highest wastewater treatment cost, and the cost penalty is about 25% of the total pollution control cost and about 7% of total process operating costs, including pollution controls.

There are, however, some identifiable impacts in regard to air pollution regulations, particularly SO_2 emissions. All of the options involve utilization or conversion of petroleum residues which contain a disproportionate part of the sulfur found in crude oil. Removal of this sulfur to limits set for SO_x emissions would have a significant impact on those options which involve the direct combustion of asphalt (vacuum bottoms), viz., options A-1 and B as described above. In the case of option A-1, the operating costs for flue gas desulfurization (FGD) accounts for more than half of the total operating cost (see Table V-4). This is why refiners prefer to comply with sulfur dioxide regulation by fuel blending rather than by installing FGD systems on process heaters and boilers. The reason FGD is not generally applied in refineries is that there are numerous small (by electric utility scale) sources within a specific refinery which require control. FGD systems for the capacities normally encountered with process heaters have high unit costs.

---

*See Appendix F, Glossary, for petroleum industry terms.

A possible alternative to options A-1 and B is to sell the asphalt to an electric utility for burning in large boilers equipped with FGD systems. This would reduce the unit cost of pollution controls due to the improved economies of scale, but would entail transportation difficulties. Furthermore, this poses a potential conflict with the Federal Energy Administration's (FEA) policies aimed at reducing the amount of petroleum used by electric utilities.

The impact of EPA regulations on the residuum upgrading processes, such as H-Oil and partial oxidation, is not unreasonable and, in most cases, the additional control required represents an incremental change to an existing sulfur recovery system. The pollution control cost for option A-3 is 25% of the total operating costs, including pollution control. Seventy-five percent of the control cost is for removal of sulfur from the flexigas and reactor gas.

EPA policies regarding sulfur emissions from refineries could certainly have an effect upon how the refiners manage their internal energy requirements. The present approach is to blend low-sulfur distillate oils with the heavy bottoms to meet an overall sulfur specification. In so doing, the yield of distillate oil product is reduced. Option A-1 presents an alternative which would make more light fuel oil available for refinery export; however, the FGD control technology available is too costly for practical implementation. Hence, the impact of EPA regulations upon the management of the refinery fuel balance in this case has an effect on the choice of options.

Option B is similarly impacted by the high cost of small source FGD. A potentially attractive modification to option B is the integration of electric power generation with process steam generation through the use of a topping cycle (steam pressure reduction with back pressure turbines). This approach could conserve total energy in the refinery in contrast with the other options which just upgrade form value. Hence, some movement in this direction may be mandated by FEA, economics notwithstanding. Asphalt combustion would be a natural choice of fuel for this scheme, were it not for the high cost of stack pollution control. Again there is a case for developing small source FGD technology, perhaps utilizing typical refinery resources such as CO, H_2, amine plants and sulfur-reduction processes.

The choice of alternates for heavy resid conversion is particularly dependent upon the type of crudes processed at a given refinery and the particular markets for asphalt, coke, and distillate fuels. Furthermore, the impact of pollution regulations is not particularly great for these options; consequently, pollution regulations are not seen as an impediment in the application of these technologies.

C. ENERGY CONSERVATION POSSIBILITIES

As a result of this assessment, two technology areas have been identified, which through added research could increase energy conservation in terms of form value availability. The first would have improving the reliability and reducing the cost of flue gas desulfurization as an objective, especially in the size range below 50 MW equivalent. The second would be concerned with developing rugged hydrocracking catalysts which can withstand the poisons normally present in petroleum residues, or alternatively, an economical residual demetallization process. The objective in this case would be to increase space velocities (more activity) and yields. Solvent de-asphalting of hydrocracker feedstocks is used for this purpose, but the energy requirement for solvent recovery is substantial. Technology improvements in these areas could increase refinery yields of clean fuels without sacrificing environmental quality.

Relative to other industrial sectors, the petroleum refining industry generally has fewer technology limitations preventing energy conservation and pollution control. At issue is not how to do the job, but rather how can the cost be borne?

III. PETROLEUM REFINING INDUSTRY OVERVIEW

A. INDUSTRY DESCRIPTION

A general description of the petroleum refining industry is presented in this section, with detail available in Appendix A. The petroleum refining industry is one of the most complex and technically sophisticated industries in the United States. There are currently about 250 U.S. refineries, ranging in size from 400,000 barrels per day (BPD) to only a few hundred BPD. Refineries vary from those which are fully integrated, highly complex plants, capable of producing a complete range of petroleum products and some petrochemicals, to very simple plants, capable of producing only a small number of products. Table III-1 gives a breakdown of plant sizes and capacities. Some refineries are modern, while others contain processing units constructed over 40 years ago. Crude slates for refineries also vary as do product lines and product properties. Because of all this variation, it is difficult to describe a "typical" refinery. Each refinery is characterized by a unique capacity, processing configuration, and product distribution.

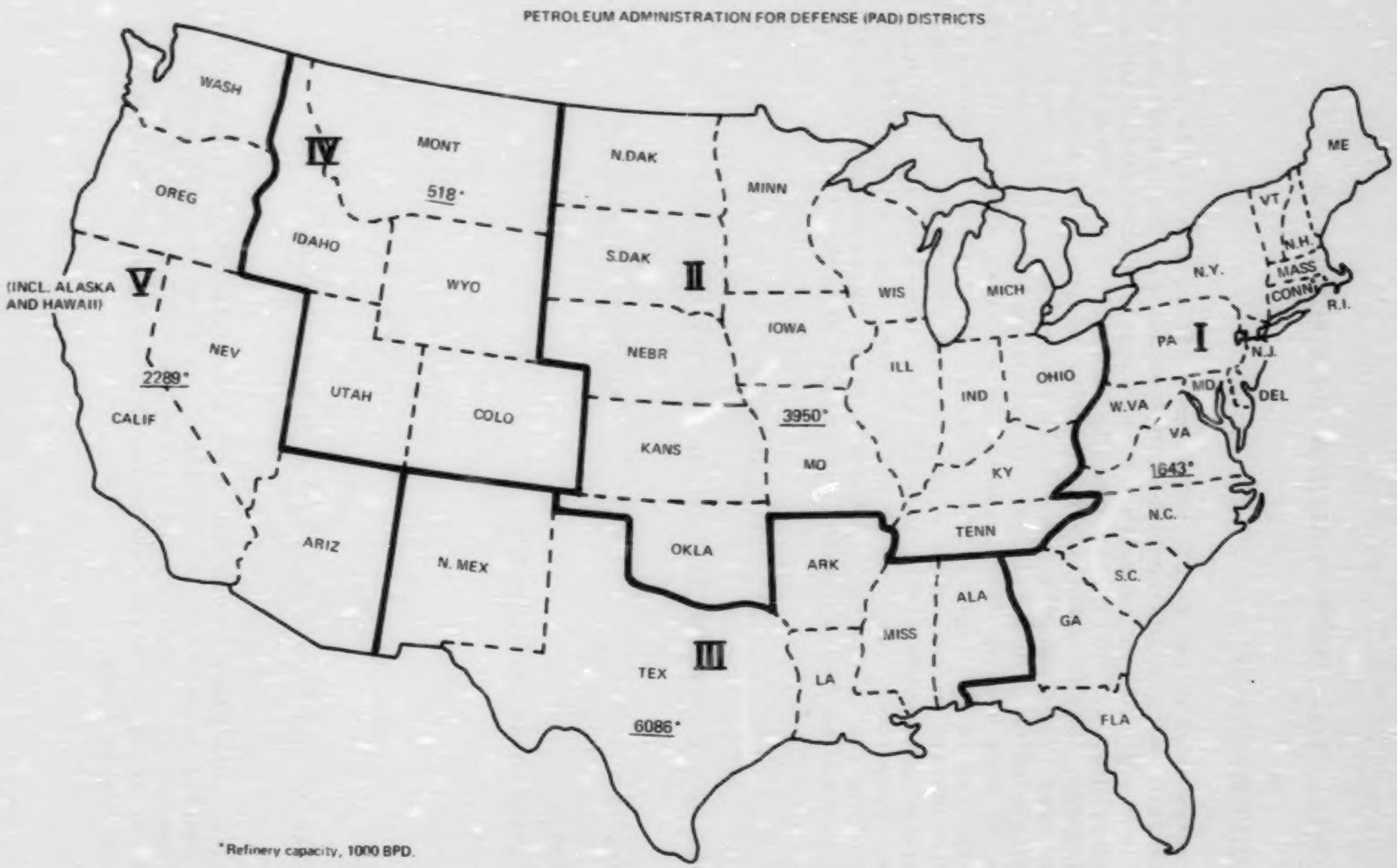
TABLE III-1

SURVEY OF U.S. REFINING INDUSTRY BY CAPACITY AND NUMBER OF PLANTS

Plant Size, BPD:	<10,000	10-100,000	>100,000
No. of Plants	74	141	44
Total Refining Capacity, 1,000 BPD	348	6,032	8,465
% of Plants	28	55	17
% of Capacity	2.3	40.7	57.0

Source: The Oil and Gas Journal, Annual Refining Survey (1975).

Although each plant is unique, there are similarities in operations based on general location of the facility. In order to collect statistics on the refining industry, the Bureau of Mines has divided the United States into five refining regions, called Petroleum Administration for Defense (PAD) Districts (see Figure III-1).


PETROLEUM ADMINISTRATION FOR DEFENSE (PAD) DISTRICTS
I
II
III
IV
V
(INCL. ALASKA AND HAWAII)
WASH
OREG
CALIF
NEV
2289*
IDAHO
MONT
518*
WYO
UTAH
COLO
ARIZ
N. MEX
N.DAK
S.DAK
NEBR
KANS
OKLA
TEX
6086*
MINN
IOWA
MO
3950*
ARK
LA
WIS
ILL
MISS
MICH
IND
KY
TENN
ALA
OHIO
GA
FLA
W.VA
VA
1643*
N.C.
S.C.
PA
N.Y.
MD
DEL
N.J.
VT
N.H.
MASS
CONN
R.I.
ME

*Refinery capacity, 1000 BPD.

Source: Bureau of Mines

Figure III-1. Petroleum Administration for Defense (PAD) Districts

Based on the Bureau of Mines figures for 1974, total U.S. refining capacity for 1974 was 14,486,000 BPD. District III has by far the greatest capacity (6,086,000 BPD). The four other districts, arranged in decreasing order of capacity, are District II (3,950,000 BPD), District V (2,289,000 BPD), District I (1,643,000 BPD), and District IV (518,000 BPD). In the period between 1960 and 1974, Districts II, III, and V experienced the greatest growth.

Small (<30,000 BPD), non-integrated refineries represent over half of the total number of refineries in the United States, while their combined capacity is less than 10% of total U.S. throughput. These facilities include operations whose primary products are materials such as asphalt and operations that are essentially topping refineries which produce a very small number of products. These refineries generally market their products nearby. Since the energy requirements and processing configurations vary widely among small refiners, and since their production comprises such a small percentage of total U.S. production, they are not considered for purposes of this study. However, because of their size, EPA regulations can place an extreme burden on small refiners and could thus alter the basic competitive structure of the industry.

B. ECONOMIC OUTLOOK

The number of refineries in the United States has decreased in the last few decades (see Table III-2), while the average size of a refinery has increased. Few new grassroot refineries have been built in the past five years; however, changes have been made in existing refineries to reflect changing technology and product demand, largely through expansion and revamping of units of existing refineries. Although there are several new facilities in the planning stages, many such projects have been either cancelled or greatly delayed primarily because of the uncertainty caused by unresolved energy and environmental issues.

Growth in petroleum product demand within the next 10 years is expected to be lower than historical growth rates, partly due to nearly zero gasoline growth (50% of total petroleum products). Companies interviewed in connection with this study agreed in general that:

1. Gasoline growth between now and 1985 will be much less than it was for the 1965-1973 period. Projections by those interviewed ranged from zero growth to 2% per year. This is attributed to improved fuel economy for automotive engines and the trend among consumers to purchase smaller automobiles.

2. Major growth areas in the next decade will be petrochemicals and fuel oils, although these represent a relatively small fraction of total petroleum products.

3. The jet fuel market will also see a slowdown in growth from that experienced between 1965-1973.

TABLE III-2

SURVEY OF OPERATING REFINERIES IN THE UNITED STATES - 1962-1972

				Charge Capacity (MMB/SD)								Production Capacity (MMB/SD)				
		Operating Refining Capacity[1]				Catalytic Cracking										
Date	Number Plants	(MMB/CD)	(MMB/SD)	Vacuum Distillation	Thermal Operation	Fresh Feed	Recycle	Catalytic Reforming	Catalytic Hydro-cracking	Catalytic Hydro-refining	Catalytic Hydro-treating	Alkyla-tion	Catalytic Polymeriza-tion*	Lube	Asphalt	Coke (MT/SD)
1/1/62	299	10.01	10.59	3.67	1.81	3.75	1.47	2.02	-	-	2.37	0.46	0.14	0.21	0.49	18.90
1/1/63	293	9.92	10.46	3.58	1.75	3.89	1.55	1.99	-	-	2.54	0.49	0.13	0.20	0.49	19.20
1/1/64	288	10.18	10.72	3.75	1.72	3.99	1.62	2.05	-	-	2.75	0.50	0.13	0.20	0.51	20.94
1/1/65	275	10.25	10.76	3.76	1.64	3.99	1.57	2.06	-	-	2.93	0.53	0.13	0.21	0.54	21.14
1/1/66	265	10.25	10.75	3.76	1.69	3.96	1.53	2.09	-	-	3.10	0.55	0.12	0.21	0.53	23.03
1/1/67	261	10.45	10.95	3.89	1.64	3.95	1.65	2.19	-	-	3.35	0.60	0.11	0.21	0.54	25.00
1/1/68	269	11.14	11.66	4.08	1.66	4.18	1.60	2.38	0.41	-	3.66	0.65	0.10	0.21	0.53	28.43
1/1/69	263	11.57	12.08	4.12	1.60	4.25	1.55	2.54	0.50	0.55	3.27	0.67	+0.25	0.20	0.57	29.43
1/1/70	262	12.15	12.65	4.55	1.64	4.37	1.49	2.78	0.60	0.54	3.51	0.75	0.29	0.21	0.58	35.49
1/1/71	253	12.68	13.28	4.74	1.56	4.51	1.46	2.89	0.73	0.54	3.81	0.78	0.31	0.22	0.60	38.77
1/1/72	247	13.09	13.71	4.85	1.53	4.57	1.26	3.17	0.84	0.63	4.26	0.82	0.29	0.22	0.62	41.47
Incremental Change																
1962-1972		3.08	3.12	1.18	(0.28)	0.82	(0.21)	1.15	0.43	0.08	1.89	0.36	-	0.01	0.13	22.57
% Crude		100.0	101.3	37.8	(9.0)	26.3	(6.7)	36.9	13.8	2.6	60.6	11.5	-	0.3	4.2	-
NPC Questionnaire & Expansion Data (MMB/CD)																
1973-1978		1.8	-	0.50	(0.18)	0.20	0.0	0.67	0.11	0.67	0.99	0.13	0.14	0.18	0.09	6.4
% Curde		100.0	-	28.0	(1.0)	11.0	0.0	37.0	6.0	37.0	55.0	7.0	8.0	1.0	5.0	-

---

[1] MMB = million barrels
CD = calendar day
SD = stream day

*Aromatic and isomerization reported beginning 1/1/69.

Source: Oil and Gas Journal, Annual Refining Reports (January 1, 1962 through January 1, 1972); NPC Refining Survey Questionnaire (1973 through 1978).

Another reason for a halt in expansion of the petroleum refining industry is the rapid increase in the construction cost of refineries. The cost, currently estimated at between $3000 and $4000 per daily barrel is up from a figure of $2000 per daily barrel of only a few years ago. The scarcity of capital is a major factor in inhibiting refining growth within the United States, particularly in light of uncertainties regarding future domestic environmental and energy policies and other governmental regulations such as divestiture.

IV. COMPARISON OF CURRENT AND ALTERNATIVE PROCESSES

A. CRITERIA FOR ASSESSING OPTIONS

We applied a subjective screening to arrive at the final selection of process alternatives for detailed analysis. The factors considered in this screening are briefly discussed in the subsections below.

1. Probability of Implementation

The timeframe of the study extends to 1990 with emphasis on the nearer term. In assessing potential process changes, the prospect of implementation in this period were considered. Among a list of applicable refinery technologies, there will be different degrees of attractiveness due to differences in commercial experience, reliability, product yields, and the like. For example, flexicoking, flue gas desulfurization, residuum hydrodesulfurization, methanol synthesis, and slurry coal partial oxidation do not connote the same degree of implementability within the period of interest. An appreciation of this variability was factored into the selection.

2. Pollution Control

Pollution regulations with which future process changes must comply include the 1983 Best Available Technology Economically Achievable (BATEA) standards for water effluents from refineries and New Source Performance Standards (NSPS) for water discharges and air emissions, especially those limiting sulfur emissions. This latter regulation affects the sulfur content of the refinery fuel which must currently comply with existing local regulations within each region. While pollution control aspects were factored into the selection process, the overriding consideration was the potential for energy conservation. However, once the final options were selected, pollution control aspects became the dominant concern.

3. Energy Considerations

The unit operations carried out within a refinery are, on balance, net consumers of energy. Since better policing and waste heat recovery were not in the scope of the study, the major consideration in the selection of energy-conserving changes was the amount of form value improvement obtained. Processes that improve the yields of LPG's, light, medium and heavy distillates in deference to atmospheric and vacuum bottoms were heavily weighted. Speculations on energy trends identified by the industry interviews were also considered. These included a feeling that on the West Coast end-use control of methane (pipeline gas) is a possibility. Another possibility is that the existing electric power rate structure which favors industrial users may be overhauled resulting in an increased electricity charge to the refiner at some unknown time in the future. Several factors, such as its hydrophilic nature, will apparently preclude the use of methanol as an additive to motor fuel.

4. Feedstock Availability

Due to declining domestic production and the reduction of Canadian imports, changes in crude slate are certain during this time period. Alternative sources of domestic and imported crude will be required to make up these deficiencies. The specific sources selected will depend on the geographical region. For example, on the West Coast, Alaskan North Slope crude will be available in this time period, while on the East Coast the major sources would be North Africa and the Middle East. The selection of process options was made within the context of this changing feedstock environment. Within the timeframe of interest, ADL does not see any significant amounts of OCS gas and oil being produced off the East Coast. Exploratory wells have yet to be drilled, and it would take 10 years before significant production levels could be achieved.

5. Market Considerations

Several market considerations were factored into our evaluation of possible options. Specifically, in regard to product demands, gasoline growth is expected to be small due to improvements in engine efficiency and decreases in automobile weight. Product demand is expected to occur mainly for the distillate and residual fuel products, e.g., jet fuels, kerosenes, diesel fuels, No. 2 and industrial fuels. Essentially all companies ADL contacted forecast zero or very small gasoline growth. Petrochemicals are expected to grow at a relatively large rate. However, the relative impact of petrochemical growth on energy consumption within refineries is expected to be small.

The future market for asphalt and coke is difficult to estimate. The West Coast industry representatives contacted expect that coke will continue to be exported. Also, industry representatives believe that more and more asphalt (resid) will be upgraded to middle distillates and gasolines. The ease of this upgrading is very crude-specific. Some refiners and districts will have great difficulty in converting asphalt (resid) to more valuable products and will have a much higher percentage of the crude barrel to convert, e.g., PAD V. In PAD I, coking is not prevalent and asphalt must be either sold or converted. If this asphalt cannot be sold and/or upgraded, one of the alternatives is to burn it as fuel. For example, SOCAL at Richmond has burning capability and has burned solvent de-asphalted bottoms (with the addition of cutter stock).

6. Process Economics

While it is difficult to generalize on refinery process economics, certain aspects were noted in our selection. For example, a potential alternative to asphalt conversion is direct combustion in process heaters and boilers. In view of the capital investment required for conversion processing, direct combustion could be a more economic alternative. Another economic factor noted was that the cost per million Btu for flue gas desulfurization and resid hydrodesulfurization is very close and the final choice depends on the conditions imposed in a specific situation.

B. SELECTION OF STUDY ALTERNATIVES

Due to the variation and complexity of the process options available for the refining of petroleum, and the uncertainty surrounding future product demands, the selection of appropriate study alternatives was an important subtask of this assessment. To ensure that reasonable alternatives were identified, we conducted a series of interviews with industry representatives. During these interviews, various potential processing changes were introduced to determine the potential for implementation. Based upon the results of these interviews, we selected three scenarios and five process options for detailed study from a list of eleven possible candidates. The complete list of options and descriptions of those which were not evaluated in full are briefly discussed in Appendix E. The options retained are discussed below.

1. Option A: Utilization of Asphalt to Increase Yields of Higher Value Products

a. Direct Combustion

To conserve higher form value products, asphalt could be used for direct combustion in process furnaces and boilers. This would displace existing imported natural gas, internally generated refinery fuel gas, and residual fuel oil (containing asphalt as well as higher valued, medium and heavy gas oil to reduce its viscosity and sulfur level). However, asphalt is generally very high in sulfur content (2-5%) and is difficult to desulfurize economically because of the high levels of asphaltenes and metals which deposit on the catalyst of the desulfurization process and make it inoperable.

Hence, the combustion of asphalt requires the availability of a reliable flue gas desulfurization (FGD) process. Several refineries now use the Wellman-Lord FGD process for SO_2 removal from Claus plant tail gas. Exxon has used a sodium-based FGD process for SO_2 removal from catalytic cracker regenerator flue gas and the process is apparently operating satisfactorily. Of course, FGD has been studied extensively for the utility industry.

It is important to note that burning high-sulfur fuel, plus FGD, is comparable in cost to indirect desulfurization of residual fuel oil and burning the lower sulfur fuel. Hence, because of the uncertain reliability of existing FGD processes, there is not much incentive for a company to install an FGD process in a refinery if lower sulfur fuel oil is available. However, the alternative of burning asphalt has not been widely investigated, but is examined in the present study.

b. Hydrocracking

Another means by which refiners will meet future demands for light, high form valued products is by direct hydrocracking of residuum. In this process, hydrogen and asphalt are contacted with a catalyst at elevated pressures and temperatures to produce refinery gases, gasoline precursors, distillates, and a byproduct low-sulfur fuel oil. Although the hydrocracking of gas oils is widely practiced, the energy/environmental consequences of upgrading residuum by this means has not been widely investigated, and is included in the present study.

Nelson (The Oil & Gas Journal, December 17, 1973, p. 69) presented a cost comparison of different routes to produce a 1.5% sulfur fuel oil and a 0.5% sulfur fuel oil. In general, according to his study, vis-breaking produced the lowest cost fuel oil, and heavy oil cracking the highest cost fuel oil.

M.W. Kellogg (The Oil & Gas Journal, May 27, 1974, p. 75) responded to Nelson's article and made the following points: "Our (MWK) studies have indicated that hydrocracking (HDC) processing for low-sulfur fuels has particular merit as compared to direct desulfurization when:

1. The feed is high in metals, i.e., greater than about 75 ppm Ni + V;

2. The sulfur specifications in fuel oil are low -- less than 0.3%;

3. It is desired to produce low-viscosity rather than high-viscosity residual fuels;

4. The high steam-producing capability of the HDC process can be integrated with overall refinery requirements in grassroot refineries. (In essence, the HDC process utilizes the poorest portion of the crudes (asphaltenes) to satisfy the major portion of the refinery steam requirements);

5. Flexibility to alter distribution towards gasoline and lighter products in place of low-sulfur fuels is desired."

The above discussion illustrates that the choice of process for an existing refinery is clearly very complex, and while the process selected by ADL may not be the best for all situations, it does represent a reasonable choice.

c. Flexicoking

Coking is a process wherein asphalt can be exposed to elevated temperatures to crack it to higher form-valued light products, with coke produced as a byproduct. The use of the process is restricted to those refiners having a market for the coke. The concept is desirable as a means to convert low-valued asphalt, and will be more widely practiced in future decades as the coke market permits.

Flexicoking is basically an extension of the fluid coking process with addition of a coke gasifier. The gasifier converts the coke formed into a clean, low-sulfur gaseous fuel. The process can convert up to 99% of a vacuum residuum into lighter (and more valuable) liquid and gas products.

Pilot plant data have been published for flexicoking of three different residua: Iranian heavy, Bachaquero, and West Texas sour (after propane de-asphalting). The published yield data are shown in Table IV-1.

The first commercial flexicoker, rated at 22,000 BPSD*, is under construction at Toa Oil Company's refinery in Kawaski, Japan, and is scheduled for startup in spring 1976. Exxon Research and Engineering Company has recently published operating data on a 750-BPSD* flexicoking pilot plant (The Oil & Gas Journal, March 10, 1975, p. 53).

Flexicoking is clearly a process which has strong potential to upgrade residuum and leave a minimum of heavy products for further disposal. Its use must be evaluated relative to alternative means of accomplishing this upgrading, including the energy/environmental consequences.

2. Option B: Internal Power Generation Using Asphalt

The existing electric power rate structure, which favors industrial users, may be overhauled as a result of consumer pressures, resulting in an increased electricity charge to the refiner. Industry representatives told ADL that if a flat rate structure is implemented (such as is advocated by Massachusetts' consumer groups), internal power generation will be economic for large (100,000 BPD) refineries. Furthermore, since there are parallel pressures to upgrade asphalt and to limit sales of high-sulfur products (such as asphalt), a case was investigated to determine the energy/environmental effects of generating power within the refinery using asphalt as an energy source. Internal generation of electrical power is currently practiced by many refiners, although only in large refineries and not with an asphalt fuel. This is technically feasible and will perhaps become economic at some time in the near future. In principle, the power plant could burn coal, oil, or asphalt. The burning of a high-sulfur fuel is predicated on the availability of an operable and economic flue gas desulfurization process.

---

*BPSD = barrels per stream day

TABLE IV-1

FLEXICOKER YIELDS FROM VARIOUS RESIDUA

Feed Properties	Iranian Heavy	Bachaquero	West Texas sour asphalt
°API Gravity	5.1	2.6	-0.2
Con. carbon, wt %	21.4	26.5	34.0
Sulfur, wt %	3.43	3.66	4.6
Nitrogen, wt %	0.77	0.81	0.65
V + Ni, ppm	525	1,040	137

	Iranian Heavy		Bachaquero		West Texas sour asphalt	
Yields	wt %	LV %	wt %	LV %	wt %	LV %
C_3 and lighter	9.9	---	10.5	---	11.3	---
C_4 =	1.3	---	1.4	---	1.5	---
iC_4	0.1	---	0.1	---	0.1	---
nC_4	0.5	---	0.6	---	0.6	---
C_5/360° F	11.0	15.4	10.3	14.7	9.2	13.4
360°/975° F	50.8	55.1	44.2	48.3	33.4	36.9
Coke	1.2	---	1.5	---	2.0	---
Coke gas		15.6*		21.1*		30.0*

---

*FOEB basis

LV - Liquid volume

Source: Oil & Gas Journal, March 10, 1975

3. Option C: Alternative Methods for Hydrogen Generation

Hydrogen is a critical feedstock for many refinery processing steps, such as hydrocracking of heavy feedstocks into higher form-valued products and desulfurization of fuel oils. We feel that the availability of hydrogen will remain critical to the industry over the next several decades.

Essentially all manufactured hydrogen within a refinery is produced by steam methane reforming. As natural gas becomes scarce - perhaps curtailed, and certainly more expensive - alternative sources of hydrogen will become more attractive. Among the alternative sources considered were:

a. Partial oxidation of residuum,

b. Coal, and

c. Naphtha and similar boiling-range materials.

Five companies were interviewed extensively to obtain their viewpoints on the likely process routes to obtain hydrogen.

All companies agreed that in the next decade or so there would be a switch away from steam methane reforming to alternative feedstocks. Presently, the preferred feedstock, based on economics, is naphtha. One company in California has completed the conversion of all its steam-methane reformers to allow use of naphtha as a feedstock. This company has retained methane as the fuel, but this may also be converted to naphtha.

Several economic studies showed that light naphtha (C5-235F) is the most economic feedstock for hydrogen production (after methane). This conclusion stems partly from the fact that naphtha of this boiling range is a "low" octane gasoline component with poor lead susceptibility. These same studies also indicated that coal is a poor feedstock for hydrogen production. The choice of other feeds is determined by the relative prices of LPG, naphtha, and resid, plus the fact that maintenance costs, in general, increase as the gravity of the feed increases.

Although naphtha is an attractive feedstock now, price differentials and availability may shift the preference toward coal or residuum over the next decade. Since residuum is available in refineries and does not exhibit the transportation/handling problems of coal, its use as a source of hydrogen is most likely.

Partial oxidation of hydrocarbon stocks to make hydrogen is a proven commercial process, and more than 100 plants are in operation using either Shell or Texaco technologies. Most installations are not in the United States, and the feedstocks used are generally lighter than residual oil. Standard Oil of California has installed a partial oxidation process at its El Segundo refinery to produce hydrogen from residual. Information available to ADL indicated that the process is not now being used, and that SOCAL makes hydrogen at El Segundo from other feedstocks. The technology to operate the plant with heavy oil is available, but operating costs are greater than for other feedstocks.

Conversion of residual oil to hydrogen is a technically feasible option, but the price of the hydrogen is, in general, not competitive. This conclusion of course is subject to the values of alternate feedstocks, and to the availability of alternate markets for residual, such as coke or heavy fuel blending. In California, considerable coking capacity has been installed and large quantities of coke are exported. Market projections indicate that coke will continue to be exported in large amounts during the 1975 to 1985 period.

C. REFINERY BASE LINE AND PROCESSING ALTERNATIVES

1. Refinery Base Line Used for Comparison

a. Refinery Simulation

To simulate existing U.S. petroleum refineries, we used three cluster models to describe the regional characteristics of the refining industry and the processing configurations typical of the regions. Each of these cluster models simulates the average operation of three similar refineries located in the same refining region.* The refining regions selected include the East Coast (PAD I), Gulf Coast (PAD III), and West Coast (PAD V). The development and calibration of these computer models was accomplished under other EPA contracts with ADL (ADL, 1975a and 1975b). The final reports are in review.

b. Validation of Model

A critical objective in developing the model was to ensure that it effectively represented "real" refineries, as well as the section of the United States containing the refineries. When the models were assembled, an extensive calibration effort was undertaken by Arthur D. Little, Inc., in collaboration with representives of the Environmental Protection Agency and an American Petroleum Institute/National Petroleum Refiners Association task force.

Data on raw material intake, fuel consumption, and product make for each of the refinery clusters and for the regions of the United States containing these clusters were obtained from the Bureau of Mines. Direct operating data on these refineries were compiled by representatives of the EPA. Processing information was obtained from petroleum literature sources and from individual industry sources. Using this processing information, we ran the individual cluster models on the computer, checked results against the refinery data, and updated them as required. We continued this task until each cluster model was calibrated with the industry data, assuring an accurate representation of the refinery clusters being simulated.

These cluster models, therefore, represent typical refineries in the specific refining regions in terms of crude type, processing configuration, and product slate, but range in crude capacity from 160,000 to 220,000 BPD. These production characteristics, along with the energy consumed by each process unit, are summarized for the calibration cases in Appendix C. Although the actual calibration runs are based upon 1973 Bureau of Mines data, we believe they are generally representative of conditions to mid-1974. The value of these models is that refinery operating characteristics, such as unit capacities, fuel consumption, and sulfur distribution, can be determined for future scenarios having different crude intakes, product makes and process configurations.

---

*A list of actual refineries simulated by the cluster models as well as typical flow sheets are provided in Appendix C.

c. Study Base Line

In assessing the impact of processing changes in the refining industry, we used a 1985 refinery configuration (existing refineries) and energy consumption base line. We selected 1985 because currently there are two broad energy-related trends at work which will significantly alter refinery operations as characterized by the 1974 calibration cases. These trends include:

- Phasing out of natural gas as a fuel in refineries, and
- Removal of lead from gasolines.

The effects of these trends are already in evidence and were factored into the calibration runs. However, by 1985 when future process changes would be fully implemented, we believe these underlying trends will have reached a steady-state condition. We forecast that this steady-state condition will find essentially no natural gas being used in refineries and a lead-free gasoline pool. Since these shifts in energy utilization are expected to occur and their effect could influence the conclusion of this assessment study, we chose a base line that characterized refinery operations in 1985 for existing refineries.

The refinery capacity and energy consumption data for 1985 operation are summarized in Appendix C. The shifts in energy utilization between 1974 and 1985 are highlighted in Table IV-2. The phaseout of natural gas by 1985 has brought about an increase in the use of blended fuel oil for refinery fuel. In addition, the total refinery fuel requirement has increased by 1985. This is due to the removal of lead from gasoline and will require the U.S. refining industry to manufacture gasoline with clear (unleaded) octane numbers about four numbers higher than when manufacturing leaded gasoline This increased octane will be achieved primarily by upgrading existing catalytic reformers to operate at higher severities and by building new high-severity reformer capacity. However, operation of catalytic reformers at high severity does not yield as much gasoline as does the present low-severity operation, so additional crude oil and fuel are consumed in making a fixed quantity of motor gasoline (ADL, 1975a).

d. Base Line Pollution Characterization

(1) Airborne Pollutants

The major airborne pollutants emitted by refineries have previously been identified by process step (EPA, 1973) and a compilation of emission factors is presented in Appendix D. The daily emission rates of the primary pollutants of concern are tabulated by process in Tables IV-3, 4, and 5 for the three base line refinery clusters. The sulfur dioxide emissions are based on refinery sulfur balances for each cluster. The emission quantities for the other pollutants were determined from the emission factors cited. Sulfur emissions from the Claus plant are based on 99.95% recovery of the sulfur in the acid gas. Vacuum ejectors are the source of hydrocarbons associated with vacuum distillation.

TABLE IV-2

ANTICIPATED SHIFTS IN REFINERY ENERGY USAGE CAUSED BY NATURAL GAS AND TEL PHASE-OUT

	East Coast		Gulf Coast		West Coast	
	1974	1985	1974	1985	1974	1985
Crude Run, 10^3 BPD	188	198	222	218	155	164
Purchased Natural Gas, FOE/day[1]	2,500	---	5,400	---	6,390	---
Refinery Fuel, FOE/day[1]						
Gas	4,810	6,990	5,450	6,880	7,190	7,500
Oil (blended)	7,210	9,640	5,350	10,700	2,140	11,690
Total Fuel, FOE/day[1]	14,520	16,630	16,200	17,580	15,720	19,190
Purchased Steam, 10^6 lb/day	5.6	5.7	0.5	0.5	0.7	0
Electricity, 10^3 kWh/day	713.7	838.7	744.2	875	766.8	897.3

[1] FOE = Fuel Oil Equivalent, 6.3×10^6 Btu.

TABLE IV-3

MAJOR AIRBORNE EMISSIONS FROM BASE LINE REFINERY

Location: East Coast
Year: 1985

Process/Pollutants	Particulates (lb/day)	SO_x (lb/day)	Hydrocarbons (lb/day)	NO_x (lb/day)	Aldehydes (lb/day)	Ammonia (lb/day)
• Combustion (heaters and boilers)	8,979	45,200	2,671	38,085	373	Negligible
- Gas-fired	881	---	1,321	10,129	132	Negligible
- Oil-fired	8,098	---	1,350	27,956	241	Negligible
• Fluid catalytic cracking	2,780	52,800	13,684	4,416	1,182	3,359
• Vacuum distillation	Negligible	Negligible	8,931	Negligible	Negligible	Negligible
• Claus plant	---	2,280	---	---	---	---
Total emissions	11,759	100,280	25,286	42,501	1,555	3,359

Source: Arthur D. Little, Inc. estimates

TABLE IV-4

MAJOR AIRBORNE EMISSIONS
FROM BASE LINE REFINERY

Location: Gulf Coast
Year: 1985

Process/Pollutants	Particulates (lb/day)	SO_x (lb/day)	Hydrocarbons (lb/day)	NO_x (lb/day)	Aldehydes (lb/day)	Ammonia (lb/day)
• Combustion (heaters and boilers)	9,855	27,600	2,798	40,999	398	Negligible
- Gas-fired	867	---	1,300	9,969	130	Negligible
- Oil-fired	8,988	---	1,498	31,030	268	Negligible
• Fluid catalytic cracking	3,585	8,400	17,644	5,694	1,524	4,331
• Vacuum distillation	Negligible	Negligible	9,854	Negligible	Negligible	Negligible
• Claus plant	---	680	---	---	---	---
Total emissions	13,440	36,680	30,296	46,693	1,922	4,331

Source: Arthur D. Little, Inc. estimates

TABLE IV-5

MAJOR AIRBORNE EMISSIONS FROM BASE LINE REFINERY

Location: West Coast
Year: 1985

Process/Pollutants	Particulates (lb/day)	SO_x (lb/day)	Hydrocarbons (lb/day)	NO_x (lb/day)	Aldehydes (lb/day)	Ammonia (lb/day)
• Combustion (heaters and boilers)	10,765	58,800	3,055	44,769	434	Negligible
- Gas-fired	945	---	1,418	10,868	142	Negligible
- Oil-fired	9,820	---	1,637	33,901	292	Negligible
• Fluid catalytic cracking	1,708	7,600	8,404	2,712	726	2,063
• Vacuum distillation	Negligible	Negligible	10,309	Negligible	Negligible	Negligible
• Claus plant	---	3,680	---	---	---	---
Total emissions	12,473	70,080	21,768	47,481	1,160	2,063

Source: Arthur D. Little, Inc. estimates

(2) Wastewater

The major sources of refinery wastewater in descending order of importance are cooling water blowdown, boiler blowdown, and process water. Many of the constituents found in petroleum appear in the wastewater, especially for those water streams which were in contact with petroleum products during processing. Constituents found in the wastewater include free oil, dissolved hydrocarbons, sulfur and nitrogen compounds, coke, and inorganic particulates. As a result, refinery wastewater has both a chemical (COD) and biochemical (BOD) oxygen demand.

Using a recent survey of refinery raw waste loads (EPA, 1973), we estimated typical constituent and total waste loadings for the reference refineries; these are shown in Table IV-6. Treated effluent waste loads meeting BATEA (1983) treatment levels are also shown in Table IV-6. A detailed discussion of refinery pollution sources and loadings is presented in Appendix D.

(3) Solid Waste

The nature and quantity of solid wastes emanating from refineries are highly variable and still the subject of investigation. Solid wastes basically fall into two main groups; namely, those that are generated continuously and those generated intermittently. Continuous wastes (disposal frequency < 2 weeks) are generated in process units and wastewater systems and include spent catalyst, coke fines, and biological sludges. Intermittent wastes are generally a result of tank and process equipment cleanings and the volume depends greatly upon the individual refinery housekeeping practices.

Using estimates of aggregated solid-waste generation in the United States, we estimated representative solid-waste loadings for the three base line refineries; they are shown in Table IV-7.

e. Summary of Pollution Control Costs

Wastewater treatment costs based upon the treatment steps described in Appendix D were determined for each base line refinery, assuming total implementation of the 1985 BATEA levels for refineries. These costs are summarized in Table IV-8. A detailed cost breakdown of the aggregate costs appearing in Table IV-8 is presented in Appendix D.

The costs shown in Table IV-8 are for treatment systems handling the entire wastewater generated by the refineries. It was necessary to use this basis in assessing the implications of anticipated process changes, because the wastewater from individual unit operations would not normally be treated separately, but rather combined with the rest of the refinery effluents.

Hence, the net effect of a process change is to add or subtract an incremental load from the basic treatment facility.

TABLE IV-6

BASE CASE REFINERY - CHARACTERISTICS OF RAW AND TREATED PROCESS WASTEWATER

Wastewater Parameter	East/Gulf Coast Refinery (200,000 BPD)				West Coast Refinery (175,000 BPD)			
	Raw Wastewater (1)		Treated Effluent (2)		Raw Wastewater (1)		Treated Effluent (2)	
	(lb/day	(mg/l)	(lb/day	(mg/l)	(lb/day	(mg/l)	(lb/day	(mg/l)
BOD_5	5,000	91	260	4.7	4,375	91	227	4.7
TSS	1,920	35	260	4.7	1,680	35	227	4.7
COD	14,000	254	1,420	25.8	12,250	254	1,243	25.8
Oil & grease	1,920	35	50	0.9	1,680	35	44	0.9
Phenolics	200	3.6	1.0	0.018	175	3.6	0.88	0.018
Ammonia (as N)	700	12.7	316	5.7	612	12.7	277	5.7
Sulfide	700	12.7	4.4	0.08	612	12.7	3.85	0.08
Total chromium	Varies		13(4)	-	Varies		11.4(4)	-
Hexavalent chromium	Varies		0.22(4)	-	Varies		0.19(4)	-
Flow Rate	6,600,000 gpd				5,780,000 gpd			

Notes: (1) Raw wastewater characteristics derived from EPA Development Document (EPA 440/1-73/014)

(2) Raw wastewater characteristics are for wastewater downstream of the plant API oil separator.

(3) Treated effluent loadings are based on the Best Available Technology Economically Achievable (BATEA) treatment level for 1983, as defined in "Effluent Guidelines and Standards - Petroleum Refining", 40 CFR 417 FR May 9, 1974.

(4) Chromium - contaminated cooling water blowdown may be discharged as a separate waste stream.

(5) Waste loads include stormwater runoff from process areas but not from offsite facilities such as tank farms.

TABLE IV-7

ESTIMATION OF BASE CASE SOLID WASTE GENERATION

Solid Waste Stream	Total Estimated U.S. Refinery Solid Waste Generation Rate[1] (Wet Basis)		Base Case Refinery Solid-Waste Generation	
			200,000 BPD East and Gulf Coast Refineries	175,000 BPD West Coast Refinery
	tpy	Tons/10^6 BPD	tpy	tpy
Neutralized HF alkylation sludge	18,430	4.25	264	231
Coke fines	3,820	0.88	54.6	47.8
FCC catalyst fines	34,100	7.85	487	426
API separator sludge	76,500	17.6	1,091	955
Non-leaded product tank storage	91,500	21.1	1,308	1,145
Slop oil emulsion solids	37,200	8.57	531	465
Cooling water sludge	41,300	9.5	589	515
Waste bio sludge[2]	89,500	20.6	1,277	1,117
Storm water runoff silt	33,200	7.65	474	415
Boiler feed water lime sludge	858,700	197.9	12,270	10,736
Kerosene filter clays	4,800	1.1	68	60
Exchanger bundle cleaning sludge	1,450	0.33	20.4	17.9
Air flotation float	66,200	15.25	946	828
Crude tank sludge	830	0.19	11.8	10.3
Cooling tower sludge	500	0.12	7.5	6.6
Leaded tank sludge	1,840	0.42	26	22.8
TOTAL	1,359,870	313.3	19,425	16,700

Notes:

1. Source: "Assessment of Industrial Hazardous Waste Practice, Petroleum Refining - Draft Report," U.S. Environmental Protection Agency, January 1975.

2. Waste bio-sludge estimate is on a wet basis, and therefore differs from the estimates of incinerated sludge presented in the water pollution section.

3. Unit loads are based on a total U.S. refining capacity of 14 x 10^6 BPD, or 4,340 x 10^6 barrels per year.

TABLE IV-8

SUMMARY OF BASE LINE WASTEWATER TREATMENT COSTS
(Total Annual Cost Including 20% Return on Investment)

	East Coast Refinery (200,000 BPD)	Gulf Coast Refinery (200,000 BPD)	West Coast Refinery (175,000 BPD)
	Base Case Refinery		
Process wastewater treatment costs	4,334	4,266	3,974
Cooling tower blowdown treatment costs	685	653	567
Total annual wastewater treatment costs	5,032	4,919	4,541

Note: Wastewater treatment costs are based on total implementation (from ground up) of the BATEA level (Best Available Technology Economically Achievable), 1983.

The situation for airborne emissions is somewhat different. The treatment systems are usually designed for specific sources and installed near the point of discharge. Furthermore, with all of the options considered, the primary environmental concern is that of sulfur emission control. For example, asphalt (vacuum resid) is a high-sulfur product which, when burned, will generate SO_2 in concentrations exceeding present combustion source standards. Similarly, if asphalt is upgraded by hydrocracking or coking, the resulting gaseous product will contain high concentrations of H_2S. In either case, these sulfur emissions represent an increase in the total sulfur emission of the refinery, since with the base line case the asphalt product is sold for non-combustion uses, i.e., roofing felt, paving materials, and so on.

Other pollutants, such as nitrogen oxides, hydrocarbons, and particulates, will remain at approximately the same emission rate level with respect to all of the technologies. In particular, we have made the following assumptions about each of these pollutants:

- Hydrocarbons - Present refinery practice is to vent safety valves and emergency vents as well as startup vents to a plant flare system, thereby minimizing the most important source of fugitive hydrocarbon emissions within the refinery. Other sources of fugitive emissions (valve stems, pump and compressor seals, and petroleum storage tanks) all require standard mechanical-type controls, such as seals, floating roof tanks, and the like. However, these sources will be the same for both the base case and the new technologies.

- Nitrogen Oxides - Nitrogen oxide emissions are a problem only with combustion sources and may differ slightly from one option to the next. This is particularly true for the combustion of West Coast asphalt which is expected to have a slightly higher fuel nitrogen content than other asphalts. But since the control techniques for nitrogen oxide are limited to combustion modifications, which are currently incorporated in new combustion installations, we judged that NO_x emissions for the base case and for the options would be about the same.

- Particulates - In the options where asphalt replaces gas as a fuel, there will be an increase in particulate emissions, but these will be controlled simultaneously with SO_2 emissions. In cases where asphalt replaces fuel oil, there is very little difference in the particulate emissions. In cases where asphalt is converted directly to higher valued products, the processes being considered do not generate particulate emissions.

Hence, the only control technologies considered for air pollution were:

1. The use of scrubbers to control flue gas SO_2 and particulates; and

2. The use of sulfur recovery processes, such as a Claus plant with tailgas cleanup, to control H_2S emissions in fuel gas streams.

Therefore, lumped pollution control costs for the entire refinery were not determined in assessing the impact on the process changes. However, base line pollution costs for those emissions affected by changes were determined for comparison purposes. These base line costs, together with the control costs for specific process changes, are presented under those sections dealing with the particular energy-conserving process change. General control costs for tailgas cleanup and flue gas desulfurization systems are presented in Appendix D. The total solid waste produced by each base line refinery is approximately as follows:

Location	Solid Waste Generated (tpy)
East	19,400
Gulf	19,400
West	16,700

Annual solid waste disposal costs, based upon a typical unit disposal cost of $5/ton, are $91,000 for the East and Gulf Coast refineries and $80,000 on the West Coast.

f. Refining Operating Cost

Typical operating costs (excluding crude costs and product values) for existing refineries in 1985 are summarized in Table IV-9. Because there is considerable uncertainty in projecting future escalation and energy costs, the operating costs are presented in constant 1975 dollars. Included in the costs are the capital and operating costs associated with fuel desulfurization and sulfur recovery to meet the existing regulations for combustion sources previously noted. Claus plant sulfur recovery is 99.95%.

2. Process Change A-1: Direct Combustion of Asphalt in Process Heaters and Boilers

This alternative was evaluated within the context of the East Coast refinery cluster model.

a. Process Description and Current Status of Commercialization

In converting process heaters from the use of refinery gas and lighter oils to asphalt, several factors must be accounted for. First, because asphalt is very viscous, it must be heated and kept warm to maintain reasonable flow rates in distribution lines around the refinery. To the extent possible, lines are recirculated to keep them open and to avoid having the asphalt "set up" in stagnant areas. Lines that are not recirculated must be heat traced. All asphalt lines must be insulated for heat conservation. Although asphalt combustion is technically feasible and the tendency is toward this direction, it is not widely practiced today, and there is yet some question about its economic attractiveness.

Figure IV-1 shows a typical flowsheet for circulating and burning asphalt in refinery process heaters.

b. Potential for Energy Conservation

This scheme, as are the others related to utilizing asphalt, is intended primarily to upgrade the overall form value of refinery products rather than actually increasing the overall thermal efficiencies within the refinery. Table IV-10 shows the overall refinery energy balance with and without direct

TABLE IV-9

EXISTING REFINERY
CAPITAL INVESTMENTS AND OPERATING COSTS
(Constant 1975 dollars)

	East Coast	Gulf Coast	West Coast
Crude Capacity 103 BPD	200	200	170
Capital Investments ($ x 10^6) (1st Qtr 1975 Basis)			
Approximate Book Value	200.0	200.0	170.0
Operating Costs ($ x 10^3/day)			
Purchased steam	21.9	1.6	---
Electricity	18.0	12.3	16.2
Cooling water	16.6	17.3	12.5
Maintenance	30.8	31.5	29.6
Manpower	53.8	44.5	71.0
Catalysts and chemicals	20.1	16.6	16.5
Fixed Costs			
Depreciation	40.3	40.3	34.3
Capital charges[1]	480.4	434.5	359.6
TOTAL	681.9	598.6	539.7
$/Bbl	3.40	3.00	3.17

---

[1] On investment and working capital.

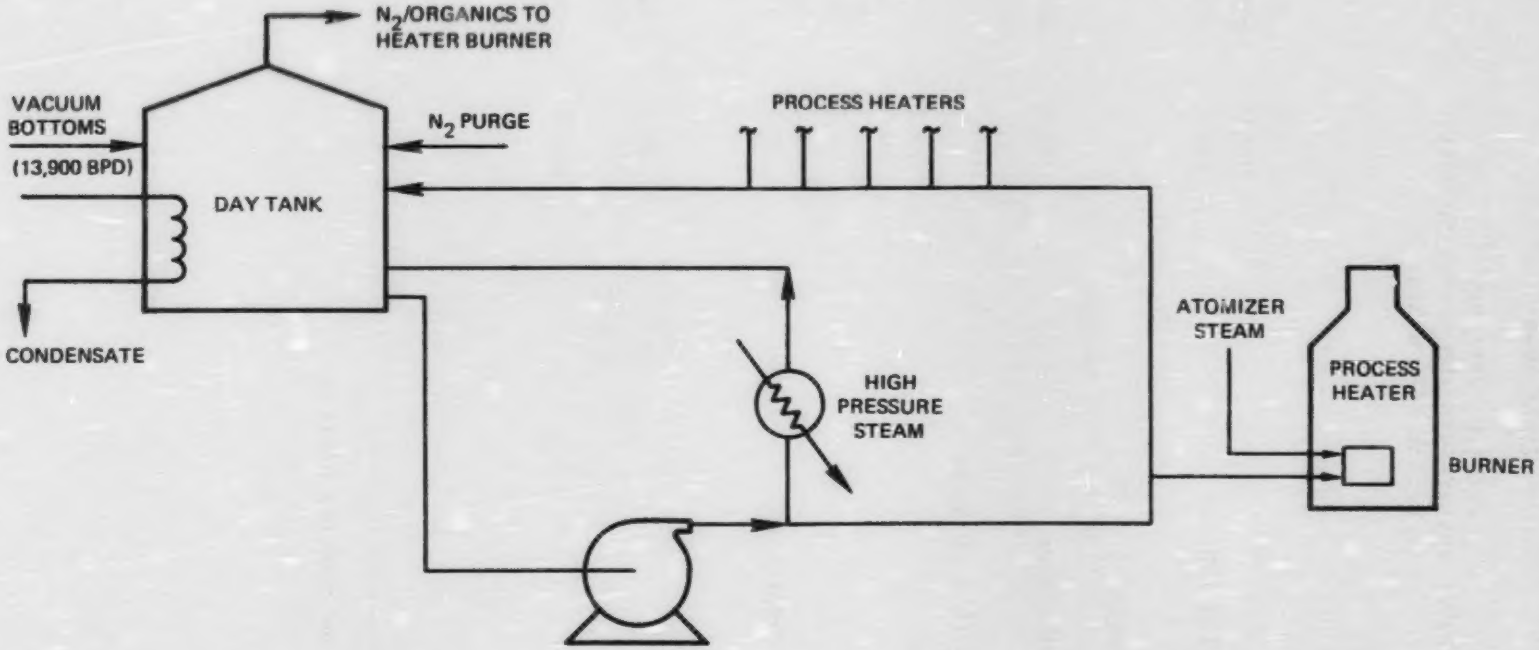

N$_2$/ORGANICS TO HEATER BURNER
VACUUM BOTTOMS
(13,900 BPD)
DAY TANK
N$_2$ PURGE
CONDENSATE
PROCESS HEATERS
HIGH PRESSURE STEAM
ATOMIZER STEAM
PROCESS HEATER
BURNER

Figure IV-1. Vacuum Residue Distribution System

TABLE IV-10

COMPARISON OF REFINERY ENERGY BALANCE WITH AND WITHOUT DIRECT COMBUSTION OF ASPHALT

Energy In	East Coast Base Line (10^9 Btu/day)	Base Line with Asphalt Combustion (10^9 Btu/day)
Crude	1,157.7	1,157.7
C_4's	8.8	8.8
Natural gasoline	32.5	32.5
FCC feed (storage)	62.2	62.2
Reformer feed	33.6	33.6
Electricity	8.9	8.9
Steam	8.2	9.7
TOTAL	1,311.9	1,313.4
Energy Out		
C_1/C_2	---	16.9
Mixed olefins	27.1	27.1
Distillate products	958.2	1,019.0
Resid, asphalt, coke	201.0	123.3
TOTAL	1,186.3	1,186.3
Input $\frac{\text{(energy out)}}{\text{(energy in)}}$ =	0.904	0.903

combustion of asphalt. The overall refinery conversion efficiency is unchanged. In this case, refinery gas, distillate, and residual oil production are all increased through internal consumption of asphalt.

c. Pollution Control

(1) Identification of Pollutants

The sources and quantities of pollutants for direct combustion of asphalt are identified in Figure IV-2 and Table IV-11. The only new source of pollution is the flue gas generated by burning asphalt.

(2) Water Pollution

It is anticipated that an East Coast refinery employing direct combustion of asphalt will have a treated process wastewater effluent and a treated cooling tower blowdown that will be of the same volume and composition as that of the base case East Coast refinery. Consequently, the wastewater treatment costs will also be the same.

Note, however, that the purpose of this option is to replace refinery fuels, so that they may be sold outside the refinery. Depending on the sulfur content of the incremental residual oil produced and the regional outlets, it is possible that SO_2 scrubbers will be required on the customer's boilers. These scrubbers would produce a wastewater and/or solid-waste stream requiring control prior to discharge.

(3) Air Pollution

This method proposes increasing the form value of refinery product outputs by burning excess asphalt (vacuum tower bottoms) rather than fuel oil or, in some cases, refinery gas. The result of this is that more fuel oil will be available for sale to outside consumers.

From an environmental point of view it is important to look at this option from two perspectives: the national emission perspective and the local refinery perspective. The implications of this option on both of these perspectives is discussed below.

(a) National emission perspective - From a total energy point of view, this option results in an increase in the amount of oil per barrel of crude which is burned in a stationary combusion source. In the base case, the refinery fuel is supplied from refinery gas and from fuel oil; the asphalt is sold as a product. In option A-1, some of the refinery gas and all of the fuel oil is displaced by asphalt. The products in this case are refinery fuel gas and fuel oil which can be sold as fuel and will eventually be burned. The difference, then, is:

(1) Although the same amounts of refinery gas and fuel oil are burned in both cases, the location of the combustion source is changed, since in the base case the combustion is concentrated at the local refinery, while with option A-1 the combustion sources are distributed throughout the marketing area of the refinery; and

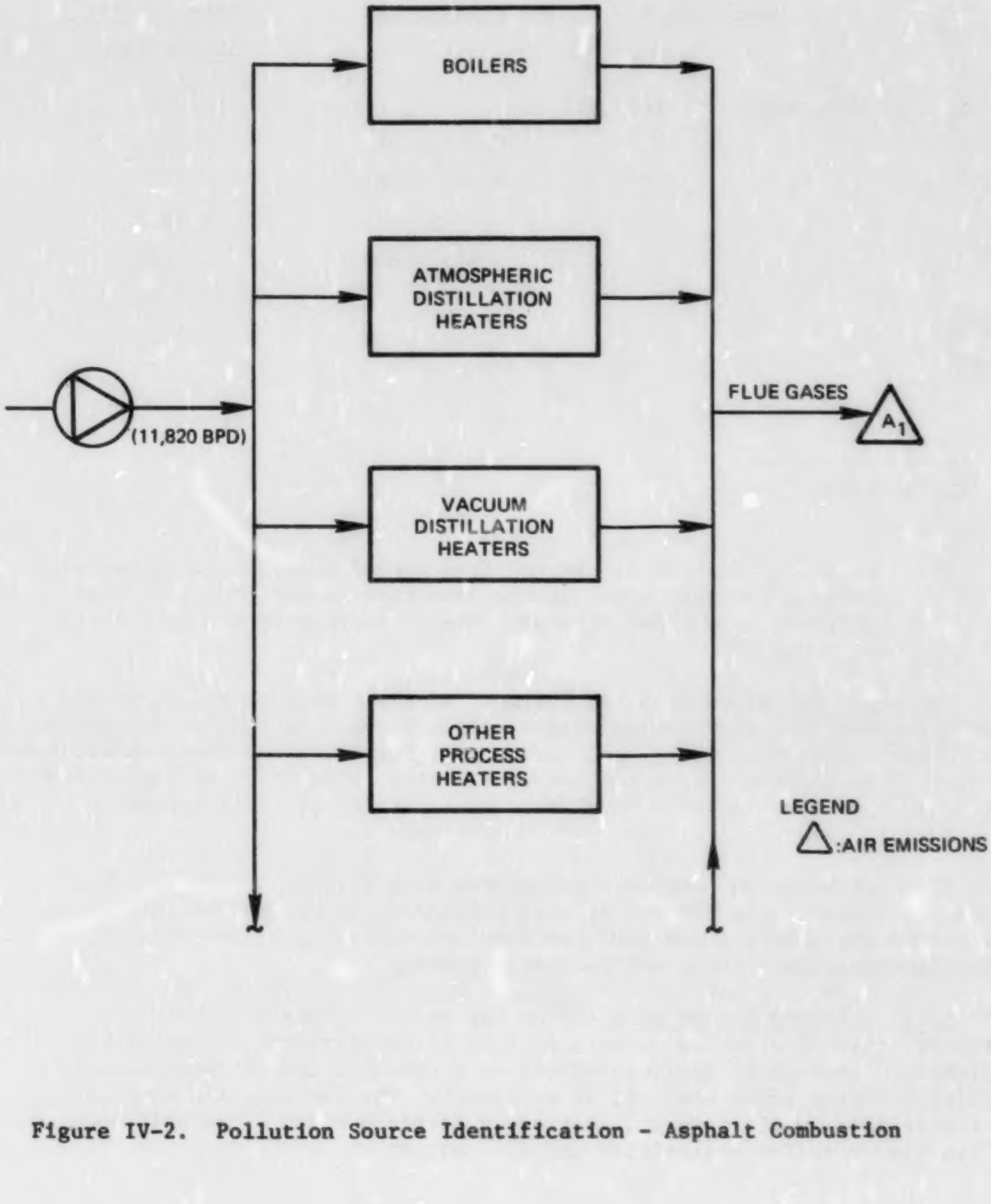

BOILERS
ATMOSPHERIC DISTILLATION HEATERS
(11,820 BPD)
FLUE GASES
A1
VACUUM DISTILLATION HEATERS
OTHER PROCESS HEATERS
LEGEND
△:AIR EMISSIONS

Figure IV-2. Pollution Source Identification - Asphalt Combustion

TABLE IV-11

ASPHALT COMBUSTION POLLUTION PROFILE
(Basis: 11,820 BPD/day asphalt burned)[1]

Stream No.	Description	Pollutant	Estimated Emission Rate (lb/hr)
			Uncontrolled
A_1	Flue gases (collective)	SO_x	15,395
		Particulate	502
		Hydrocarbons	83
		NO_x	1,729
		Aldehydes	15

---

[1] 12,330 FOED

(2) The use of asphalt is changed from one of construction materials (paving, roofing, etc.) in the base case to one of a fuel (for boilers and process heaters), thereby increasing the amount of oil burned.

The total emissions from the refinery will not only change significantly because the uncontrolled emission from asphalt combustion is greater than emissions from refinery fuel combustion, but emissions due to petroleum combustion by the refinery's customers will also increase. The impact that this will have on a region's total emissions, of course, will depend on what fuel is used in lieu of the added available oil.

The most important emission associated with this fuel replacement is oxides of sulfur. The control of this pollutant can be achieved either at the combustion source using flue gas desulfurization or at the refinery using hydrodesulfurization and fuel oil blending.

(b) Local refinery perspective - From the point of view of the local refinery, this alternative amounts to form value upgrading through fuel switching. Instead of selling asphalt as a product, the refinery will be selling a higher grade fuel oil as a product. The uncontrolled emissions at the refinery will, however, increase with the burning of asphalt because of its higher sulfur composition and ash content.

Under present operating practice at refineries, this option will result in a significant increase in the amount of pollution control required. At the present time refineries blend fuel oil with other low-sulfur feedstocks so that the blended fuel used in the refinery remains within acceptable limits for combustion without stack gas sulfur controls. This practice may come under increasing government pressure to sell rather than burn higher form value products.

(c) Control technology - The appropriate control technology for this option is flue gas desulfurization (FGD) for asphalt combustion sources. The base line comparison is hydrodesulfurization and blending (indirect desulfurization) of fuel oil.

Specific technological considerations relating to these different control technologies have been discussed many times in previous EPA reports and will not be elaborated on in detail here. Examples of the cost of control of these alternatives for a 250 million Btu/hr source are shown in Table IV-12. It is important to point out that control costs depend on the size of the combustion unit being controlled. But in a given refinery the number or size of process heaters for steam boilers is widely variable and often smaller than the example shown. For these small sources, the capital investments for FGD are highly speculative. However, the estimate is sufficient for showing the relative differences in the two cases being considered. In general, the unit cost will be higher for smaller sources.

Generally, fuel oil used within the refinery is a blend of both high- and low-sulfur feedstocks, such that the resulting mixture has a sulfur content within the limits allowed for uncontrolled stationary combustion sources. As an indicator of the implied environmental costs of this strategy, we have modeled this strategy (somewhat simplistically) as follows: a portion of each barrel of high-sulfur residual fuel oil is desulfurized so that, when recombined, the blended oil will meet local standards. For example, consider a 2% sulfur residual fuel oil available in a refinery, which can be desulfurized to approximately 0.3% sulfur. According to the blending model, approximately 82% of the oil would have to be desulfurized before the final blend could meet a standard of approximately 0.6% sulfur. If hydrodesulfurization costs approximately \$3.00/bbl, this amounts to a net cost of 39¢/10^6 Btu of fuel heating value. As can be seen from Table IV-12, the cost for the hydrodesulfurization/blending option is less than the cost for flue gas desulfurization. Clearly, these averages are only illustrative and do not represent all refinery situations, but they do tend to show the rationale behind the blending strategy adopted by most refineries throughout the country.

That being the case, one is inclined to believe that a more logical strategy might be to retain the fuel oil and refinery gas within the refinery as fuels and to export asphalt to large power plants. Although asphalt is more difficult to burn than residual oil because of the greater viscosity, it is easier to burn than coal, for example.

TABLE IV-12

EXAMPLE OF SO_2 CONTROL FOR PROCESS HEATER OR BOILER[1]

Basis: 250 x 10^6 Btu/hr (heat input), 2% Sulfur

	Flue Gas Desulfurization
Total capital cost, $000	2,161
Operating cost, $000/yr	
Variable costs:	
Labor, 1 man/shift (incl. supv. & overhead) @ $14.85	130
Maintenance, @ 5% of direct capital	83
Utilities	
- Electric power, 240 kWh/10^6 scf @ $0.03/kWh	189
- Water, 50 gpm @ $0.35/$10^3$ gal	9
- Fuel, 2.5 x 10^6 Btu/ton of S @ $2.17/$10^6$ Btu	7
Total variable costs	418
Fixed costs:	
Depreciation, 16 years	135
Insurance & taxes, @ 2% of capital	43
Total fixed costs	178
Total production cost	596
Credit for Sulfuric Acid, @ $25/ton (net)	(93)
Return on investment, @ 20% of capital	432
Total Annual Cost, $000/yr	935
Unit cost, $/$10^6$ Btu	$0.47/$10^6$ Btu
Comparative cost for hydrodesulfurization/blending, 0.82 bbl/bbl @ $3.00/bbl	$0.39/$10^6$ Btu

---

[1] Based on MgO scrubbing, sulfuric acid recovery.

It is unlikely that utilities will rapidly (or willingly) change over to firing a difficult fuel, like asphalt, particularly when the Government policy is to switch electric utilities onto coal. A more probable strategy for utilizing the heavier constituents of crude is through further processing of the bottoms, as suggested in the other options relating to this strategy.

(4) Solid Waste

Implementation of direct combustion of asphalt will not appreciably alter the quantity or characteristics of the base case East Coast refinery total solid-waste stream.

A sulfuric acid product is produced from the FGD systems, part of which might be used to offset acid requirements for alkylation, depending on acid purity. Alternatively the acid would presumably be marketed; however, no credit or disposal cost was assigned to this byproduct.

d. Cost Factors

The conversion of process heaters from refinery fuel to asphalt would entail considerable expense relative to the initial investment in heaters. Not only would a fuel day tank and fuel pumps be required, but heaters and traced and insulated piping would also be required. Further, steam would be required for atomization to ensure clean burning of the fuel. Finally, the burners themselves would have to be replaced by new combination fuel burners, a costly investment.

The total cost, including pollution controls, to convert to asphalt combustion on the East Coast is shown in Table IV-13. The cost of controlling SO_x emissions is clearly dominant.

e. Technical Considerations

Several technical factors would affect the decision to go to direct combustion of asphalt. Since this material contains more sulfur and ash than other fractions, and these will generally be high enough to require emission controls, the economics of installing the controls on these dispersed, relatively small heaters may be prohibitive. We set a criterion of a firing rate of 35×10^6 Btu/hr, below which asphalt would not be substituted in a process heater, to avoid the need for small, dedicated pollution control devices. By this method more than 98% of the refinery fuel would be asphalt in a typical 1985 East Coast refinery.

Another consideration was that the nitrogen content of asphalt is higher than that of natural gas and light oils, and combustion may result in increased NO_x emissions. Injection of steam or other means to moderate the flame temperatures may be required. As mentioned earlier, atomizing steam injection at a rate of 0.1-0.3 lb steam/lb fuel will be required for atomization to ensure clean burning with a minimum of soot formation.

TABLE IV-13

PRODUCTION COSTS: OPTION A-1 (DIRECT COMBUSTION)

Product: Process Heat

Daily Capacity: 77,680 x 10^6 Btu

Process: Direct Combustion

Fixed Investment: $1,800,000

Stream Days/Yr.: 330

Location: East Coast

Depreciation Period (yr): 16

Year Used for Costing Purposes: 1975

	Units Used in Costing or Annual Cost Basis	$/Unit	Annual Cost (10^3)
VARIABLE COSTS			
Raw Materials			
Byproduct Credits			
Energy			
• Purchased Fuel			
• Purchased Steam	338 x 10^9 lb	3.85/10^3 lb	1,300.0
• Electric Power Purchased	8000 kWh	$0.03/kWh	0.2
• Miscellaneous			
Energy Credits	--		
Water			
• Process (Consumption)	--		
• Cooling (Circulating rate)	--		
Direct Operating Labor (Wages)	3% of CAP		54
Direct Supervisory Wages			
Maintenance Labor			
Maintenance Supervision			
Maintenance Materials and Supplies			
Labor Overhead			
Misc. Variable Costs/Credits[a]			
Royalty Payments	--		
FIXED COSTS			
Plant Overhead			
Local Taxes and Insurance 2%			36
Depreciation 16 yrs. S.L.			112.5
TOTAL PRODUCTION COSTS			1,502.7
Return on Investment (pretax) 20%			360.0
POLLUTION CONTROL (Incremental)			2,051.0
TOTAL			3,913.7

[a] e.g., miscellaneous chemicals, catalysts, supplies, services.

In general, the turndown ratios of gas-fired burners are about 10-15:1 while for heavy oils they may be only 4-5:1. For most refinery equipment, which will usually run at or near full capacity, this should not prove a serious drawback, but it may impact on the refiner's decision to convert.

For flexibility, we have assumed that, given the option, the refiner would install combination burners, i.e., those which would be capable of burning gas or oil. This has been incorporated into the economics of Section C-2d.

The need for recirculation and asphalt heating have already been discussed in Section C-2a.

f. Effect on Intermediate and Final Products

This option is intended primarily to free up the refinery gas and blended fuel oil utilized as refinery fuel by substituting vacuum residue as illustrated earlier in Table IV-10. Therefore, incremental increases in products of higher form value are obtained while another of low form value is consumed rather than exported.

3. Process Change A-2: Hydrocracking of Heavy Bottoms

This alternative was evaluated within the context of the West Coast refining region.

a. Description and Current Status

The H-oil process has been chosen to exemplify heavy-ends hydrocracking processes. Figure IV-3 is a schematic of the process. Hydrogen and heavy oils are reacted in a novel reactor system - the ebullated bed. In this three-phase system, solid catalyst is maintained in continuous random motion by upflow of the liquid phase and hydrogen. Catalyst can be added to and withdrawn from the unit without shutting it down. The overhead vapors are quenched with direct water injection, which precludes the formation of plugs of sulfate salts or polysulfides. The liquid products are then fractionated to provide mostly intermediates for further processing. A small purge is drawn from the recycle hydrogen to prevent the buildup of inerts. This stream goes to amine scrubbing to remove H_2S and is burned as refinery fuel.

The H-oil process has been implemented at four refineries worldwide, with capacity totaling 69,000 BPD. However, an explosion in an H-oil unit at the Exxon Bayway refinery has temporarily halted the construction of these units until the cause is determined. Nevertheless, it is a good example of resid hydroprocessing technology. There are other resid hydrotreating/hydrocracking processes available (Exxon - Residfining, Chevron - Isomax), all of which could not be examined in this study because of time limitations.

In evaluating the incremental emissions from this process change, a feed consisting of asphalt product (2,070 BPD) and atmospheric bottoms (9,240 BPD) which otherwise would be blended into low-sulfur product fuel oil, was chosen. In addition, 900 BPD of asphalt feed must be burned as refinery fuel to displace refinery gas, which is required for producing 12.5 x 10^6 scf/day of hydrogen by steam reforming.

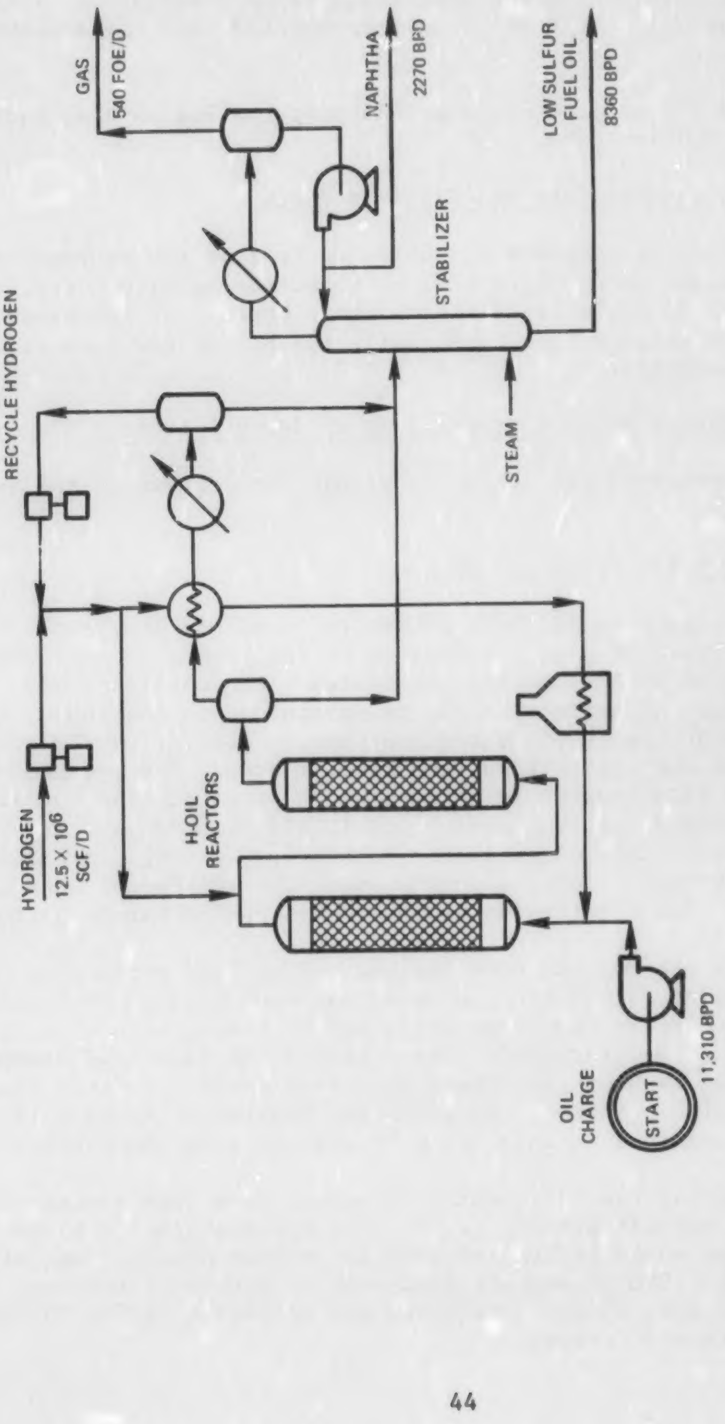

GAS
540 FOE/D
NAPHTHA
2270 BPD
LOW SULFUR
FUEL OIL
8360 BPD
STABILIZER
STEAM
RECYCLE HYDROGEN
HYDROGEN
12.5 X 10^6
SCF/D
H-OIL
REACTORS
OIL
CHARGE
START
11,310 BPD

Figure IV-3. H-Oil

b. Potential for Energy Conservation

The process has value for converting heavy ends to lighter fuels. The net make of refinery gas is about 100 FOE/day since the fuel requirement for high purity hydrogen production is more than half the heating value of both the methane and ethane produced. However, the LPG and C_5-650°F products are increased by 540 BPD and 5,330 BPD, respectively. The overall thermal efficiency (useful Btu out/Btu in) for the H-oil process is 92%. Most of the loss in thermal efficiency is due to the hydrogen production necessary for H-oil, which amounts to 1,200 scf/bbl of intake, or 12.5×10^6 scf/day. This corresponds to a 24% increase in hydrogen generating capacity above that for the base line refinery. The effect on the thermal efficiency of the refinery is to lower it slightly, as shown by Table IV-14.

c. Pollution Control

(1) Identification of Pollutants

The major emission sources associated with Option A-2 are indicated in Figure IV-4 and identified in Table IV-15. Two wastewater streams are generated. One is created by the quench water injected into the vapor stream downstream of the reactors and contains small quantities of H_2S, NH_3, and hydrogen cyanide. A second wastewater stream - not indicated on Figure IV-4--is cooling tower blowdown, approximately 57,000 gpd (1.5% of circulation rate).

The major solid emission is 3,100 lb/day of spent cobalt-molybdate-on-alumina catalyst (containing 20% heavy metals and 20% petroleum coke).

There are three sources of air emissions. Two are solely emissions of SO_x from burning the recycle hydrogen purge gas (purge is 1% of recycle rate) and from burning asphalt to displace refinery gas for hydrogen production. The remaining potential emission is hydrogen sulfide from the acid gas removal system used to treat the product gas made. This stream will ultimately be controlled with a Claus plus tailgas cleanup system. All the emission rates listed represent additions to the base line emissions created by incorporating heavy-ends hydrocracking.

(2) Water Pollution

It is anticipated that a West Coast refinery employing heavy-bottoms hydrocracking will have essentially the same wastewater volume and characteristics as the base case West Coast refinery. The only incremental (over the base case) waste load is a small sour water (sulfide-containing) stream which increases the total process wastewater flow rate from 5.78 to 5.795 mgd.

Due to its high concentration of hydrogen sulfide and ammonia, the sour water stream must be subjected to stripping prior to discharge into the central wastewater treatment facility. A certain amount of ammonia and sulfide removal can be expected in the central treatment facility. However, there will be an incremental residual, and it is therefore reasonable to expect that both the ammonia and sulfide concentration of the total treated effluent will be slightly higher than that of the base case.

TABLE IV-14

COMPARISON OF REFINERY ENERGY BALANCE WITH AND WITHOUT HEAVY-ENDS HYDROCRACKING

	West Coast Base Line (10^9 Btu/day)	Base Line with Heavy-Ends Hydrocracking (10^9 Btu/day)
Energy In		
Crude	960.6	960.6
C_4's	3.2	3.2
Natural gasoline	5.0	5.0
Fluid cat. cracker feed	20.2	20.2
Refinery feed	10.6	10.6
Electricity	9.5	10.8
Steam	---	---
Total	1,009.1	1,010.4
Energy Out		
Refinery gas	---	0.6
Mixed olefins	7.6	12.0
Distillates	682.6	734.2
Resid, vac. bottoms, tar, coke	259.6	197.8
Total	949.8	944.6
$\frac{\text{(Energy out)}}{\text{(Energy in)}} =$	0.941	0.935

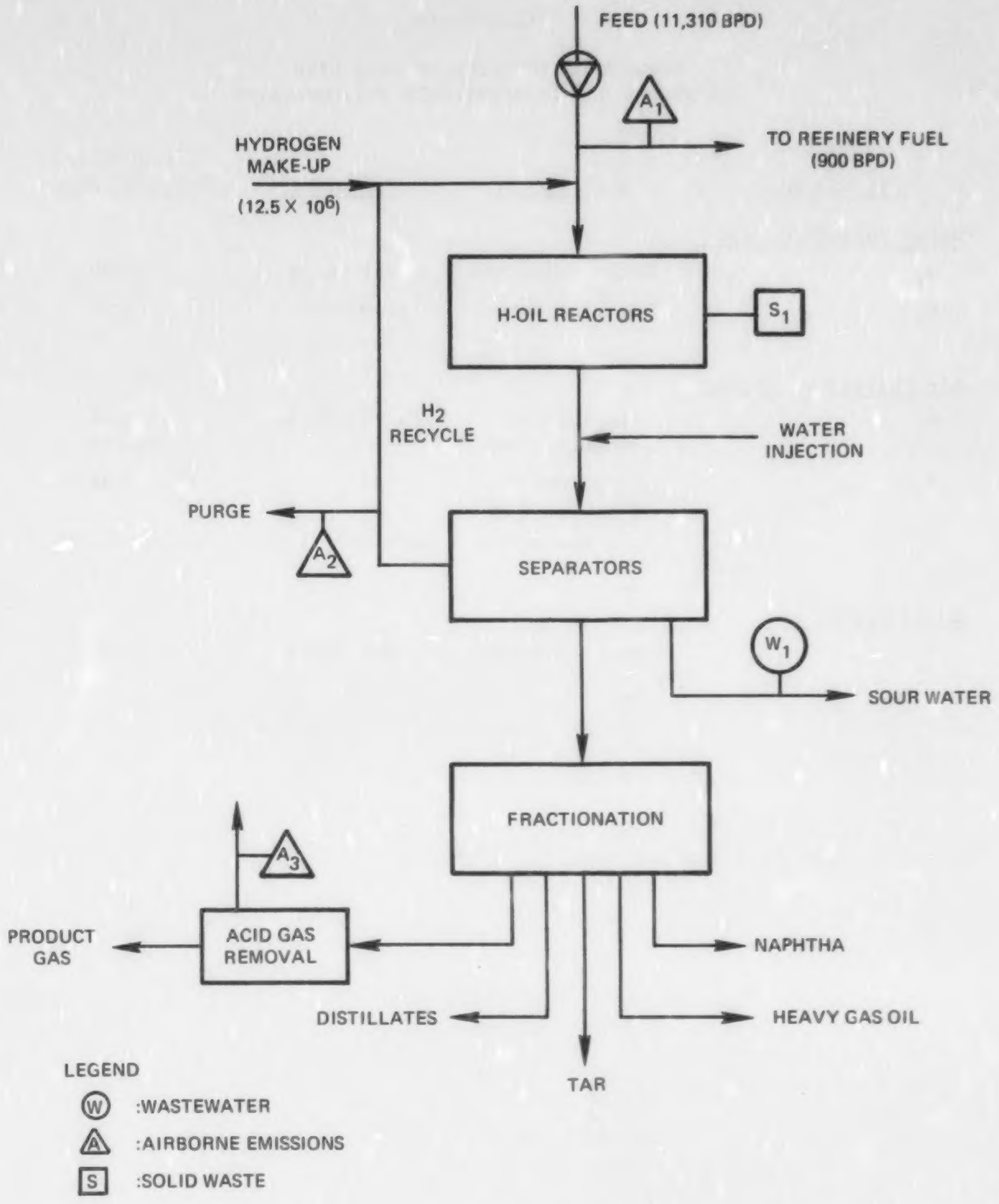

FEED (11,310 BPD)
A1
TO REFINERY FUEL (900 BPD)
HYDROGEN MAKE-UP
(12.5 X 10⁶)
H-OIL REACTORS
S1
H2 RECYCLE
WATER INJECTION
PURGE
A2
SEPARATORS
W1
SOUR WATER
FRACTIONATION
A3
PRODUCT GAS
ACID GAS REMOVAL
NAPHTHA
DISTILLATES
HEAVY GAS OIL
TAR
LEGEND
W :WASTEWATER
A :AIRBORNE EMISSIONS
S :SOLID WASTE

Figure IV-4. Identification of Pollution Sources from Hydrocracking

TABLE IV-15

SUMMARY OF ADDITION TO BASE LINE EMISSIONS DUE TO HEAVY-ENDS HYDROCRACKING

Stream No.	Description	Pollutant	Uncontrolled Emission Rate
Water Pollution, gpd			
W_1	Water injection	Sour water	14,980
W_2	Cooling water	Blowdown	57,000
Air Pollution, lb/day			
A_1	Asphalt to refinery fuel	Particulates SO_x	674 12,748
A_2	Purge gas to refinery fuel	SO_x	715 lb/day
A_3	Acid gas stream	Combined sulfur	54,530 lb/day
Solid Waste, tpd			
S_1	Spent catalyst	Wet solid	1.6

The volume and composition of the cooling tower blowdown is expected to be nearly identical to that of the base case.

Process wastewater treatment cost estimates are presented in Table IV-16. The total yearly process wastewater treatment cost is only 1.8% greater than that of the base case ($4,046,000/yr versus $3,974,000/yr).

In terms of both treated effluent waste loads and treatment costs, the water pollution impact of implementing this alternative is negligible.

(3) Air Pollution

In this option, the asphalt is used as a feedstock for a hydrocracking process such as H-oil or Isomax in which the heavy bottoms are converted to lighter fuel oils and gaseous products. The pollutants of concern in this option are H_2S generated in the gaseous products and SO_2 generated from the combustion of a small amount of asphalt. The gases from the H-oil reactor will pass through an amine scrubbing system to remove H_2S and CO_2. These are sent to a Claus plant for sulfur control. Since a refinery normally would have a Claus plant, the cost of control for this option is only the incremental cost required to expand the Claus plant to handle the additional sulfur load. In Table IV-17, the costs of this option are compared to the costs for the base case. Note that by 1985 a sulfur recovery of approximatley 99.9% will be required by most states and probably federal regulation as well. The hydrocracking case results in a net increase in sulfur of approximately 20% and a corresponding increase [illegible] sulfur control costs of approximately 12%.

In addition to the sulfur control costs for hydrotreating, the equivalent of 900 BPD of asphalt also has to be burned in the process heaters to free up refinery gas which can be used to generate the additional hydrogen needed for the hydrotreater. Since the asphalt has a sulfur content higher than the maximum uncontrolled sulfur content of a fuel, flue gas desulfurization will be required. If all of the asphalt were burned in a single process heater, the size would be approximately 240 x 10^6 Btu/hr and the cost of control would be comparable to those shown previously in Table IV-12. Since the actual distribution of the process equipment accepting asphalt as a fuel is not known, we estimated the SO_2 control costs at approximately 50¢/10^6 Btu. As can be seen from Table IV-17, the estimated cost for flue gas desulfurization is about the same as the cost for sulfur control for the hydrocracking step, even though the former controls only one-tenth as much sulfur. The major reason for this disproportionality is that the removal of acid gas benefits from the established economy of scale inherent in incremental increases in Claus plant capacity; whereas flue gas desulfurization, particularly on a scale as small as the one considered here, cannot take advantage of any economy of scale. Therefore, the total cost of controlling sulfur for this option is twice the base case.

The conclusions one must draw from the above analysis are similar to those for Option A-1, i.e., flue gas desulfurization is a relatively expensive alternative when applied to small-scale combustion sources. Consequently, low-sulfur, blended fuel oil should be used in place of refinery gas since considerably lower environmental control costs are incurred.

TABLE IV-16

ALTERNATIVE A-2

WEST COAST REFINERY WITH HEAVY-BOTTOMS HYDROCRACKING PROCESS WASTEWATER TREATMENT COSTS

Basis: 175,000 BPD, 330 days/year

Capital Investment: $10,680,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
Variable Cost			
Operating labor (including supervision and overhead)	16060 m-hr/yr	$17.15/m-hr	275,000
Maintenance (including labor and overhead)			427,200
Chemicals			
• Lime	795 tpy	$32.50/ton	24,700
• Phosphoric acid	22 tpy	$440/ton	9,700
• Chlorine	79 tpy	$140/ton	11,100
• Coagulant aid	31,800 lb/yr	$1.00/lb	31,800
• Replacement carbon	108,300 lb/yr	$0.40/lb	43,300
Fuel (gas)	6,500 MM Btu/yr	$1.31/Btu x 10^6	8,500
Steam	14,100 M lb/yr	$4.20/lb x 10^3	59,200
Electrical power	2,649,000 kWh/yr	$0.0211/kWh	55,900
Wastewater treatment sludge Disposal			
• Incinerator ash	224 tpy	$5.00/ton	1,100
• Stretford purge water	5456 tpy	$15.00/ton	81,800
Total variable cost			$1,029,300
Fixed Costs			
Depreciation @ 6.25%			667,500
Taxes & insurance @ 2%			213,600
Total fixed cost			$881,100
Total annual cost			$1,910,400
Return on investment			$2,136,000
Total			$4,046,400

Notes:

1) Capital investment adjusted to the 1975 level (ENR Construction Cost Index - 2126).

2) Wastewater treatment costs are based on total implementation (from ground up) of the BATEA level (Best Available Technology Economically Achievable), 1983.

3) Cost estimates are for the specific examples, and are in no way intended to represent industry-wide treatment costs.

TABLE IV-17

SULFUR CONTROL COSTS FOR BASE LINE AND WITH HYDROCRACKING (OPTION A-2)

	Base Line	**with Hydrocracking**
Inlet Sulfur Load, long ton per day	159	182
Capital Cost, \$000		
Claus plant	3,500	3,850
Tailgas cleanup	2,500	2,650
Total	6,000	6,500
Annual Operating Cost, \$000/yr		
Variable costs:		
Labor		
- Direct (1 man/shift @ \$6.75/hr)	59	59
- Supervision @ 10% of direct labor	6	6
Labor overhead @ 40% of labor	26	26
Maintenance @ 5% of capital	300	325
Utilities		
- Electric power @ 140 kWh/long ton, \$.0211/kWh	156	178
- Fuel @ 0.8×10^6 Btu/long ton, $\$2.41/10^6$ Btu	101	116
- Cooling water @ \$2.50/long ton in tailgas	46	52
Total variable costs	699	762
Fixed costs, \$000/yr:		
Plant overhead @ 100% of labor	65	65
Depreciation, 16 years	375	406
Insurance & taxes @ 2% of capital	120	130
Total fixed costs	560	601
Total Production Cost, \$000/yr	1,259	1,363
Credit for recovered sulfur @ \$25/ton (net)	(1,312)	(1,502)
Return on investment @ 20% of capital	1,200	1,300
Total Annual Cost, \$000/yr	1,147	1,161
Estimated incremental cost for flue gas desulfurization @ $\$0.50/10^6$ Btu	---	936
Total Sulfur Control Cost	1,147	2,097

(4) Solid Waste

Implementation of heavy-bottoms hydrocracking will slightly alter the total quantity of refinery solid waste. This process will require the disposal of a chrome-moly catalyst.

It is estimated that the waste chrome-moly catalyst will be generated at a rate of 480 tpy, thus increasing the base case total refinery solid-waste stream from 16,700 to 17,180 tpy, or 3%.

d. Cost Factors

The additional costs associated with heavy-ends hydrocracking are displayed in Table IV-18. The total - $\$6.9 \times 10^6$/yr or \$20,700/day - corresponds to $\$0.32/10^6$ Btu of product (\$2.00/FOE). No sulfur credit was taken for 25.9 tons of elemental sulfur recovered. The costs developed in Table IV-18 include facilities for increased hydrogen production.

The capital investment needed for this option is \$14.7 million, of which hydrocracking accounts for \$11.1 million.

e. Technical Considerations

It was assumed, for this case, that some asphalt could be burned in process heaters, thus freeing refinery gas for hydrogen manufacture.

Some resid hydrocracking processes incorporate a solvent de-asphalting step prior to hydrocracking to remove heavy metals, thus prolonging catalyst life. This process step has not been investigated here, but it would most likely decrease only the solid catalyst purge. Additional water treatment would be necessary since steam distillation is usually required for solvent recovery after the de-asphalting step.

f. Effect on Intermediate and Final Products

Based on 10,410 BPD to the H-oil unit, the yield structure is as shown in Table IV-19. Most of the C_1/C_2 is consumed for process heat, but all other products, except the tar, are available for further upgrading in the refinery. A comparison of the base line product slate with the product slate, including heavy-ends hydrocracking, is given in Table IV-20.

4. Process Change A-3: Flexicoking

a. Process Description and Current Status

This option was assessed within the confines of the East Coast cluster which has a high yield of asphalt and no markets for coke. Flexicoking is the combination of fluid coking with coke gasification. Although fluid coking is a commercially available technology, there are no commercially operating flexicokers. However, the process has been selected for installation by a Japanese refiner and a process development unit has been successfully operated by Exxon.

TABLE IV-18

PRODUCTION COSTS: OPTION A-2 (RESID HYDROCRACKING)

Product: Naphtha, Distillate, LPG, Tar
Byproducts: Sulfur
Annual/Capacity: 11,310 BPD

Process: Resid Hydrocracking
Batch/Continuous: Continuous
Fixed Investment: \$14.0 x 10^6
Stream Days/Yr.: 330

Location: West Coast
Depreciation Period (yr): 16
Year Used for Costing Purposes: 1975

	Units Used in Costing or Annual Cost Basis	$/Unit	Annual Cost (10^3)
VARIABLE COSTS			
Raw Materials			
• Atm. Bottoms, BPD	9,240		
• Vac. Bottoms, BPD	2,070		
• Catalyst and Chemicals		\$1,310/day	432.3
Byproduct Credits			
Energy			
• Purchased Fuel			
• Purchased Steam (10^3 lb/day)	116	\$4.20/10^3 lb	160.8
• Electric Power Purchased (kWh/day)	129,000	\$0.0211	898.2
• Miscellaneous			
Energy Credits			
Water			
• Process (Consumption) 10^3 gal/day	15	\$0.35/10^3 gal	1.7
• Cooling (circulating rate) gal/day	3,780	\$0.044/10^3 gal	54.9
Direct Operating Labor (Wages)	3 men/shift	\$6.75/hr	177.4
Direct Superv. Wages 10% of Dir. Labor			17.7
Maintenance Labor		\$1,210/day	399.3
Maintenance Supervision			
Maintenance Materials and Supplies			
Labor Overhead, 40% of Dir. Labor			78.0
Misc. Variable Costs/Credits[a]			
Royalty Payments	N.A.		
FIXED COSTS			
Plant Overhead, 100% of Dir. Labor			195.1
Local Taxes and Insurance, 2% of Cap. Inv.			280.0
Depreciation 6.25%/year			875.0
TOTAL PRODUCTION COSTS			3,570.4
Return on Investment (pretax) 20% of Cap. Inv.			2,800.0
POLLUTION CONTROL (Incremental)			1,026.0
TOTAL			7,396.4

[a] e.g., miscellaneous chemicals, catalysts, supplies, services.

TABLE IV-19

PRODUCTS FROM HEAVY-ENDS HYDROCRACKING

	BPD
Feed	
Asphalt	2,070
Atmospheric	8,340
Total	10,410
Product	
C_1/C_2	410 (FOE)
C_3/C_4	540
C_5-180°F	540
180-400°F	1,730
400-650°F	3,060
650 + gas oil	3,820
Tar	1,480
Total	11,170 (excluding C_1/C_2)

$$\text{Net volume increase} = \frac{11{,}170 - 10{,}410}{10{,}410} \times 100 = 7.3\%$$

TABLE IV-20

COMPARISON OF PRODUCT SLATES FOR THE WEST COAST BASE LINE WITH THE INCORPORATION OF HEAVY-ENDS HYDROCRACKING (1000 BPD)

	West Coast Base Line (1985)	Base Line with Hydrocracking
LPG	0.04	0.6
High-Btu gas	-	0.1 (FOE)
Unleaded gasoline	70.3	70.3
Naphtha	3.9	3.9
No. 2 distillate	22.7	22.7
Jet fuel	20.4	20.4
Kerosene	0.2	0.2
Residual fuel oil	31.3	22.1
Asphalt	2.1	-
Naphtha intermediate	-	2.3
Gas oil intermediate	-	6.9
Tar	-	1.5
Lube stocks	0.4	0.4
Coke	7.8	7.8
BTX	4.0	4.0
Petrochemicals feed	1.3	1.3
Total	164.4	164.5

Figure IV-5 is a block flow diagram of the flexicoking process. Atmospheric and vacuum bottoms can be converted to high-Btu reactor gas, low-Btu gas from the coke (flexigas), coker naphtha, and gas oil. Essentially three units comprise the flexicoker - a reactor, a coke heater-devolatizer, and a coke gasifier - all of which are operated as fluidized beds.

The resid feed is injected into the reactor, where it is thermally cracked to vapor products and coke. The vapors are quenched and scrubbed overhead to remove entrained coke. The lighter fractions proceed overhead to a fractionating section.

Coke from the reactor circulates to the heater, where it is devolatilized to yield hydrocarbon gas and residual coke. The residual coke is sent to a gasifier where greater than 90% is gasified at elevated temperatures with steam and air.

The gasifier offgas is directed back to the heater-devolatilizer for recovery of heat and then is cooled further before fines removal. The gas passes to a Stretford unit for sulfur recovery.

The flexigas has a low-Btu value (~100-130 Btu/scf), and is used to displace high-Btu refinery gas from process heaters. The reactor gas from the fluid coker is a high-Btu gas which could be sold after sulfur removal.

A small coke purge is withdrawn from the system, containing about 2% of the feed sulfur and 99% of the feed metals. The combined solids from the coke purge and fines removal amount to 2% (by weight) of the fresh feed.

The entire East Coast base line asphalt product - 18,700 BPD - was routed to the flexicoker for conversion to light liquids and gas. As shown in Figure IV-5, the products are 2,315 FOE/day of low-Btu flexigas, 2,077 FOE/day reactor gas, 348 BPD of LPG, 2,111 BPD of coker naphtha, and 8,843 BPD of coker oil (360°-950°F). The coker naphtha and gas oil are desulfurized to 10 ppm and 0.2 wt. %, respectively; gas oil desulfurization yields another 97 FOE of fuel gas. This fuel gas is combined with the reactor gas for H_2S removal.

b. Potential for Energy Conservation

A major contribution of the Flexicoking option lies in freeing high-Btu refinery gas for higher priority uses. In the case chosen here (see Figure IV-5), a net of 4,298 FOE/day, or about 27 million scf/day, of high-Btu gas are available for export. This total is due to roughly equal contributions from high-Btu reactor gas and from refinery gas displaced by coke gas (flexigas).

A second contribution to energy conservation lies in the conversion of the asphalt feed to naphtha and gas oil intermediates which become available for refining. The total yield of these distillates is 10,857 BPD, about 5% of the crude feed to the refinery. This represents a portion of crude which otherwise would not be a saleable fuel product.

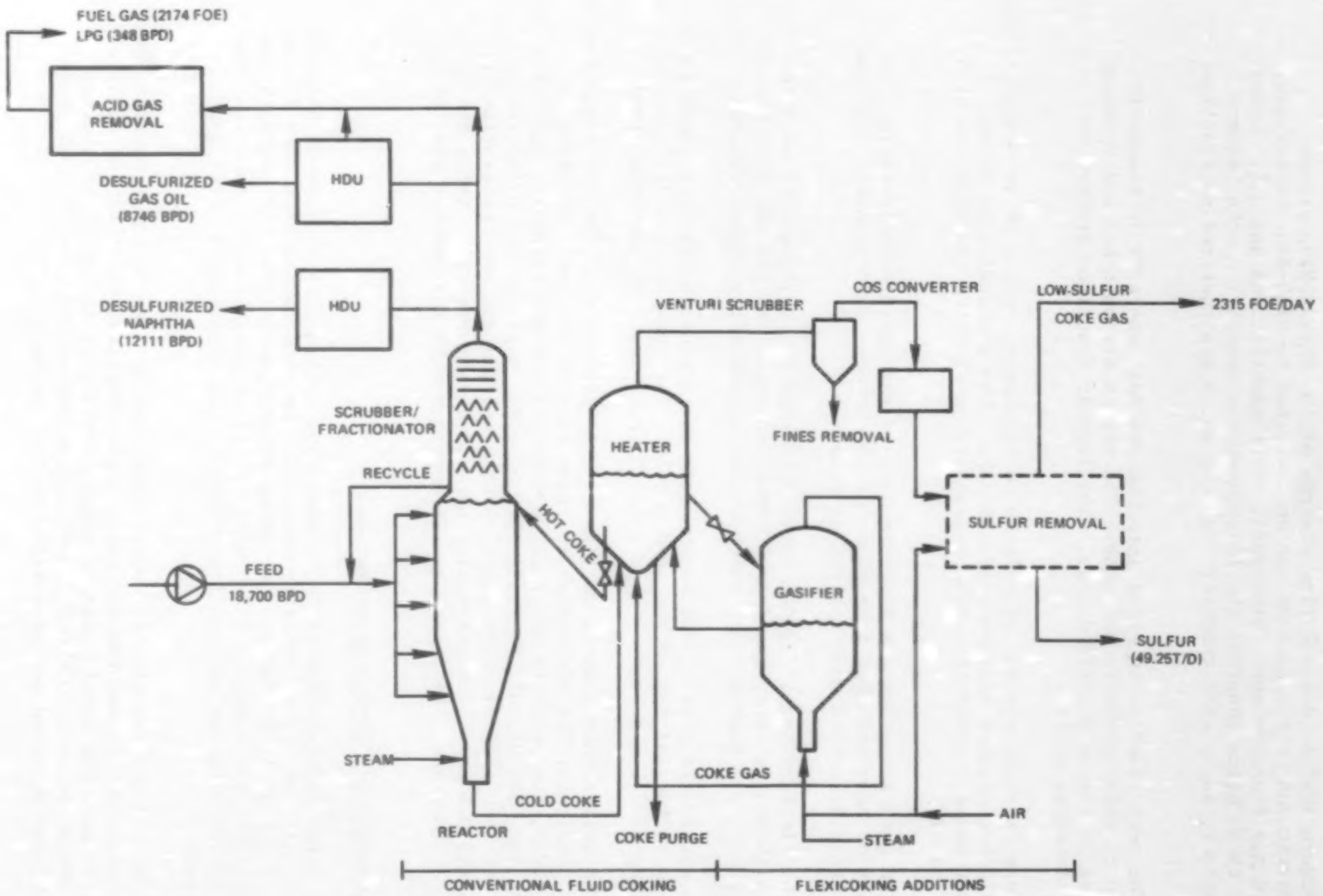

FUEL GAS (2174 FOE)
LPG (348 BPD)
ACID GAS REMOVAL
HDU
DESULFURIZED GAS OIL (8746 BPD)
HDU
DESULFURIZED NAPHTHA (12111 BPD)
SCRUBBER/ FRACTIONATOR
RECYCLE
FEED
18,700 BPD
STEAM
REACTOR
COLD COKE
HOT COKE
HEATER
COKE PURGE
VENTURI SCRUBBER
FINES REMOVAL
COS CONVERTER
GASIFIER
COKE GAS
STEAM
AIR
SULFUR REMOVAL
LOW-SULFUR COKE GAS
2315 FOE/DAY
SULFUR (49.25T/D)
CONVENTIONAL FLUID COKING
FLEXICOKING ADDITIONS

Figure IV-5. Flow Diagram and Volume Flows for Flexicoking

The thermal efficiency of the Flexicoking process (net Btu out/Btu in), excluding the heating values of the coke purge, coke fines, and product sulfur, is 96%. The effect of the low thermal efficiency is to lower the refinery efficiency, as indicated in Table IV-21.

c. Pollution Control

(1) Identification of Pollutants

Table IV-22 summarizes the incremental refinery emissions due to Flexicoking. Three wastewater streams are involved: 1) a purge from the Stretford unit, 2) sour water from steam injected into the fluid coker to strip coke particles of residual oil before they are withdrawn, and 3) 216,000 gpd of additional cooling tower blowdown (not indicated on Figure IV-6). A typical analysis of the sour water is shown in Table IV-23.

The solid coke stream, which consists of coke fines removed from the coke gas and the coke purge from the coke heater section containing 99% of the feed metals, totals 133,600 lb/day.

Airborne emissions are produced by the burning of flexigas, which was substituted for refinery gas. The emission of SO_x is in compliance with NSPS of 0.2 lb $SO_x/10^6$ Btu by virtue of the Stretford operating conditions. Due to the lower flame temperature of low Btu gas, NO_x emissions are lower than the refinery gas emissions, and Exxon claims complete removal of particulates. The acid gas removed from the coker gas (C_4 -gas) represents a gaseous pollution stream which requires treatment in a Claus/Beavon/Stretford system.

(2) Water Pollution

It is anticipated that an East Coast refinery employing Flexicoking will have a treated process wastewater stream very nearly the same as that of the base case East Coast refinery. The total waste load will be somewhat increased due to the presence of additional sour water, Stretford process purge water, and increased cooling tower blowdown.

The additional sour water stream will introduce increased quantities of hydrogen sulfide and ammonia into the total wastewater stream. The sour water stream must be subjected to steam stripping prior to discharge into the central wastewater treatment facility. The additional sour water stream will increase the process wastewater flow rate from 6.6 to 6.654 mgd.

Although there is a certain amount of removal in the central wastewater treatment facility, a sulfide and ammonia residual will result, thereby slightly increasing the sulfide and ammonia load over the base case.

The Stretford purge stream associated with the Flexicoking alternative will be increased by almost a factor of 10 over that of the base case. Since it is envisioned that the Stretford purge will be disposed of on land (via chemical fixation), this increase constitutes an increased solid-waste disposal problem rather than a water pollution problem.

TABLE IV-21

SUMMARY OF REFINERY ENERGY BALANCE WITH AND WITHOUT FLEXICOKING

Energy In	Energy Factor 10^6 Btu/Bbl	East Coast Base line 10^9 Btu/day	Baseline with Flexicoking 10^9 Btu/day
Crude	5.85	1,157.7	1,157.7
C_4's	6.3	8.8	8.8
Natural gasoline	5.6	32.5	32.5
Fluid cat cracker feed	5.6	62.2	62.2
Refinery feed	5.6	33.6	33.6
Electricity	10,500 Btu/kWh	8.9	9.6
Steam	1.45×10^6 Btu/10^3 lb	8.2	7.6
		1,311.9	1,312.0
Energy Out			
Refinery gas	6.3	-	27.1
Mixed olefins	6.3	27.1	29.6
Distillates	5.6	958.2	1,019.2
Resid, vac. bottoms	6.3	201.0	83.2
		1,186.3	1,159.1

$\frac{\text{(Energy out)}}{\text{(Energy in)}}$ =		0.904	0.883

TABLE IV-22

SUMMARY OF ADDITIONS TO BASE LINE EMISSIONS DUE TO FLEXICOKING

Stream No.	Description	Pollutant	Uncontrolled Emission Rate
Water Pollution, gpd			
W_1	Stretford purge		21,670
W_2	Coker steam injection	Sour water	54,000
W_3	Cooling water	Blowdown	216,000
Air Pollution, lb/day			
A_1	Combustion of flexigas	SO_x	2,920
A_2	Acid gas stream	Combined sulfur	159,580
Solid Waste, tpd			
S_1	Coke fines	Dry solid	66.8
S_2	High metal coke purge	Dry solid	
S_3	New Stretford unit	Elemental sulfur	49.2

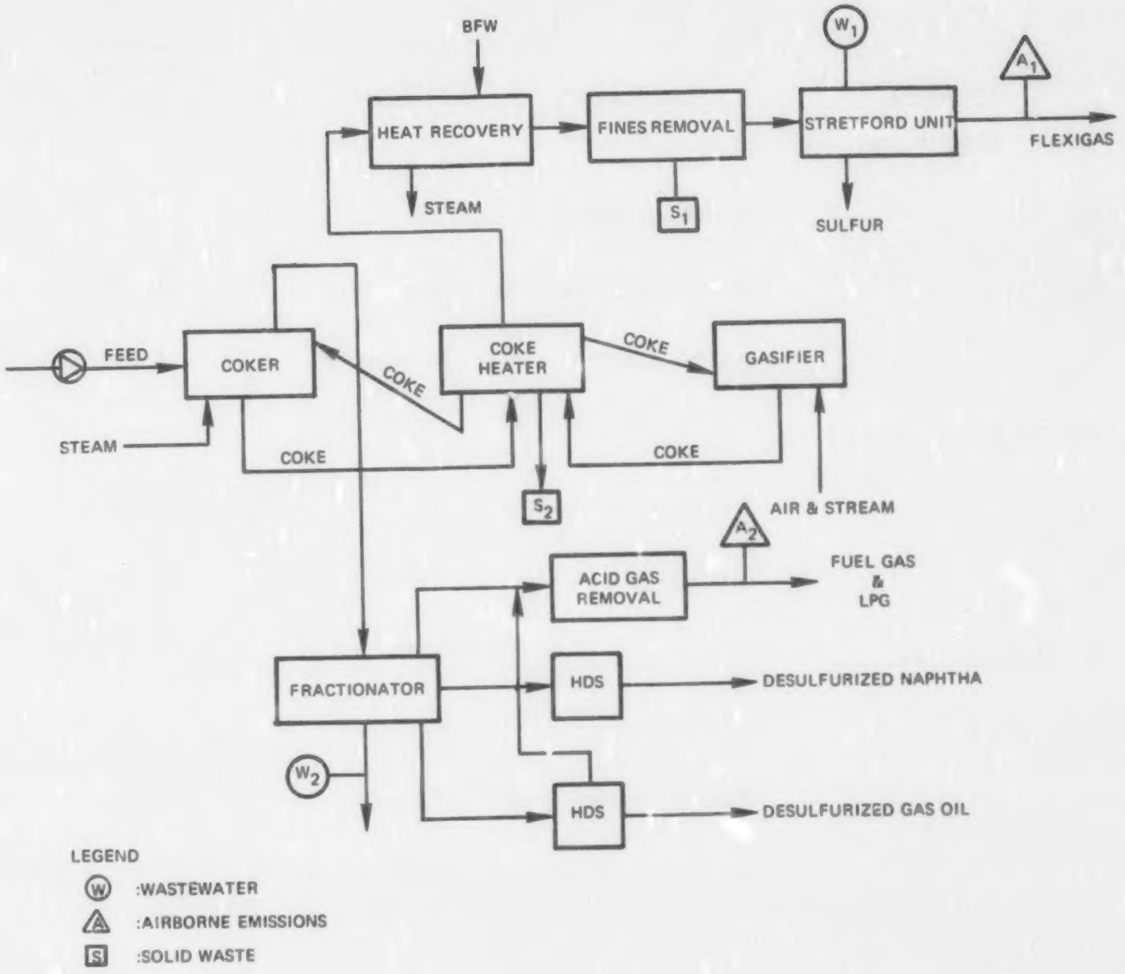

BFW
HEAT RECOVERY
STEAM
FINES REMOVAL
S_1
W_1
STRETFORD UNIT
SULFUR
A_1
FLEXIGAS
FEED
COKER
STEAM
COKE
COKE
COKE HEATER
COKE
GASIFIER
COKE
S_2
AIR & STREAM
A_2
ACID GAS REMOVAL
FUEL GAS & LPG
FRACTIONATOR
HDS
DESULFURIZED NAPHTHA
W_2
HDS
DESULFURIZED GAS OIL
LEGEND
W :WASTEWATER
A :AIRBORNE EMISSIONS
S :SOLID WASTE

Figure IV-6. Block Flow Diagram for Flexicoking and Identification of Pollution Sources from Flexicoking

TABLE IV-23

WATER ANALYSIS
FLEXICOKING SOUR WATER

Contaminants	Concentration (mg/1)
NH_3	1,000
H_2S	1,500
pH	8.5
CN (free)	0.3
SCN	9
S_2O_3	300
Suspended solids	3
Phenols	20-30
Fe	1
Ni	1
COD (by perm.)	775
V	nil

Source: Exxon R&E

The cooling water blowdown stream is increased over that of the base case by 4% (5.88 mgd versus 5.66 mgd).

All of the increases in wastewater flow rate result in a slight increase in treatment cost. It is estimated that the total treatment cost (process wastewater treatment plus cooling tower blowdown treatment) will be increased by 13% over that of the base case ($\$5.663 \times 10^6$/yr versus $\$5.001 \times 10^6$/yr). Treatment costs are presented in Tables IV-24 and IV-25. The costs are dependent on the type of prime mover used with the Flexicoking process. The cost increase would be less with electric motors or extraction turbines since less cooling water would be needed.

In terms of both treated effluent waste loads and wastewater treatment cost, the Flexicoking alternative does not present major water pollution implications.

(3) Air Pollution

The major air pollution problem associated with the Flexicoking process is in controlling sulfur in the following streams: a) fuel gas - light hydrocarbons containing H_2S; and b) flexigas - a low-Btu fuel gas containing N_2, CO, CO_2, H_2 and S.

TABLE IV-24

ALTERNATIVE A-3

EAST COAST REFINERY WITH FLEXICOKING PROCESS WASTEWATER TREATMENT COSTS

(Basis: 200,000 BPD, 330 days/year)

Capital Investment: $11,700,000

	Yearly Quantity	Cost per Unit Quantity	Yearly Cost ($/Yr)
Variable Costs			
Operating labor (including supervision overhead)	17,520 m-h/yr	$17.15/m-h	300,000
Maintenance (including labor and materials)			468,000
Chemicals			
• Lime	909 tpy	$32.50/ton	29,500
• Phosphoric acid	26 tpy	$440/ton	11,400
• Chlorine	90 tpy	$140/ton	12,600
• Coagulant aid	36,400 lb/yr	$1.00/lb	36,400
• Replacement carbon	123,750 lb/yr	$0.40/lb	49,500
Fuel (gas)	7,425 x 10^6 Btu/yr	$1.35/$10^6$	10,000
Steam	50,670 x 10^3 lb/yr	$3.85/$10^3$	195,100
Electrical power	3,026,400 kWh/yr	$0.030/kWh	90,800
Wastewater treatment sludge disposal			
• Incinerator ash	255 tpy	$5.00/ton	1,300
• Stretford Purge water	33,750 tpy	$15.00/ton	506,300
Total variable cost			$1,710,900
Fixed Cost			
Depreciation @ 6.25%			731,300
Taxes & insurance @ 2%			234,000
Total fixed cost			$965,300
Total annual cost			$2,676,200
Return on investment			$2,340,000
Total			$5,016,200

Notes:

1) Capital investment adjusted to the 1975 level (ENR Construction Cost Index - 2126).

2) Wastewater treatment costs are based on total implementation (from ground up) of the BATEA level (Best Available Technology Economically Achievable), 1983.

3) Cost estimates are for the specific examples, and are in no way intended to represent industry-wide treatment costs.

TABLE IV-25

ALTERNATIVE A-3

EAST COAST REFINERY WITH FLEXICOKING
COOLING TOWER BLOWDOWN WASTEWATER TREATMENT COSTS
(Basis: 200,000 BPD, 330 days/year)

Capital Investment: $1,417,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/Yr)
Variable Costs			
Operating labor (including super. vision & overhead)	2200 m-hr/yr	$17.15/m-hr	37,700
Maintenance (including labor & materials)			56,700
Chemicals			
• Sulfur dioxide	490 tpy	$340/ton	166,600
• Lime	522 tpy	$32.50/ton	17,000
• Sulfuric acid	245 tpy	$51.15/ton	12,500
Fuel	-	-	-
Electrical power	122,800 kWh/yr	$0.03/kWh	3,700
Wastewater treatment sludge disposal	4,790 tpy (@10% solids)	$5.00/ton	24,000
Total variable cost			$318,200
Fixed Costs			
Depreciation @ 6.25%			88,600
Taxes & insurance @ 2%			28,300
Total fixed cost			$116,900
Total annual cost			$435,100
Return on investment @ 20%			$283,400
Total			$718,500

Notes: 1) Capital investment adjusted to the 1975 level (ENR Construction Cost Index - 2126).

2) Wastewater treatment costs are based on a chrome reduction/precipitation system consisting of reaction vessels, clarifiers, chemical feed system and controls.

3) Chromium concentration in untreated cooling tower blowdown is assumed to be 30 mg/l.

4) Cost estimates are for separate treatment of chromium-contaminated cooling tower blowdown. Not all refineries treat cooling tower blowdown separately, and the amount of chromate used varies considerably.

The sulfur in the high-Btu fuel gas is removed using an amine scrubbing system, and the exhaust of that scrubbing system is sent to the refinery Claus plant. The hydrogen sulfide in the low-Btu flexigas is too low in concentration to be economically scrubbed out, and therefore this process comes with an integral Stretford unit for sulfur removal. The sulfur concentration in the product flexigas is approximately 170 ppm, which is within allowable standards for combustion without sulfur control. The cost for the Stretford unit is usually included in the cost of the Flexicoking unit. We have, however, shown estimated costs for the Stretford unit in Table IV-26 as part of pollution control, for purposes of comparison with the base case.

(4) Solid Waste

Implementation of the Flexicoking process will result in a significant increase in total refinery solid-waste generation.

The volume of the Stretford purge stream will be greatly increased over that of the base case. In addition, significant quantities of waste coke will be produced. An estimation of these increases is shown below:

	Solid Waste Generation (tpy)
Base case total refinery solid-waste stream	19,425
Increase in Stretford purge stream	28,500
Waste coke	20,700
Total	68,625

Thus, the toal refinery solid-waste volume, and hence its disposal cost at $230,000/yr, is more than three times that of the base case.

d. Cost Factors

Table IV-27 provides a breakdown of the incremental costs associated with incorporating Flexicoking into the East Coast base line - naphtha and gas oil desulfurization and 95% sulfur recovery are included. The total - $10.0 x 10^6/yr or $30,350/day - corresponds to approximately $0.33/10^6 Btu of Flexicoker products. Since the gas, oil, and naphtha are intermediates, an additional cost for upgrading would be incurred. (Note that no credit was taken for 121 tons of recovered elemental sulfur.)

The required capital investment is $22.8 million; $16.4 million of this is for the Flexicoker section alone.

TABLE IV-26

SULFUR CONTROL COSTS FOR FLEXICOKING (OPTION A-3)
(Basis: East Coast, 1985)

	Baseline	Reactor Gas Claus	Flexigas Stretford
Inlet Sulfur Load, long ton per day	99	168	42
Capital Cost, $1000's			
Claus plant	2,600	3,650	--
Tailgas cleanup	2,100	3,550	3,800
Total	4,700	6,200	3,800
Annual Operating Cost, $000/yr			
Variable costs:			
Labor			
- Direct @ $6.75/hr	59	59	19
- Supervision @ 10% of direct	6	6	2
Labor overhead @ 40% of labor	26	26	8
Maintenance @ 5% of capital	235	310	190
Utilities			
- Electric power @ $0.03/kWh	137	233	499
- Fuel @ $2.17/$10^6$ Btu	56	96	122
- Cooling water @ $0.044/$10^3$ gal	29	49	-
Chemicals @ $2.50/ton of S in tailgas	5	7	34
Total variable costs	553	786	874
Fixed costs:			
Plant overhead @ 100% of labor	65	65	21
Depreciation, 16 years	294	388	238
Insurance & taxes @ 2% of capital	94	124	76
Total fixed costs	453	577	335
Total Production Cost	1,006	1,363	1,209
Return on Investment @ 20% of Capital	940	1,240	760
Less Sulfur Credit @ $25/ton (Net)	(816)	(1,386)	(346)
Total Annual Cost, $000/yr	1,130	1,217	1,623

TABLE IV-27

PRODUCTION COSTS: OPTION A-3 (FLEXICOKING)

Product: High-Btu Gas, Low-Btu Gas, Naphtha, Gas Oil
Annual/Capacity: 18,700 BPD
Process: Flexicoking (A-3)
Batch/Continuous: Continuous
Fixed Investment: $18.0 Million
Stream Days/Yr.: 330
Location: East Coast
Depreciation Period (yr): 16
Year Used for Costing Purposes: 1975

	Units Used in Costing or Annual Cost Basis	$/Unit	Annual Cost $10^3
VARIABLE COSTS			
Raw Materials			
• Asphalt, BPD	18,700		
• Catalyst and Chemicals		$215/day	71.0
Energy			
• Purchased Fuel			
• Purchased Steam			
• Electric Power Purchased	49,780 kWh/day	$0.03/kWh	492.8
• Misc.			
Energy Credits			
• Steam, 10^3 lb/day	451		
Water			
• Boiler Feed Water, (10^3 gpd)	673	$0.60/10^3 gal	133.3
• Cooling (Circulating rate)	14,430x10^3 gal/day	$0.044/10^3 gal	209.5
Direct Operating Labor (Wages)	6 men/shift	$6.75/hr	354.8
Direct Supervisory Wages	10% of Dir. Lab.		35.5
Maintenance Labor	$1,280/day		422.4
Maintenance Supervision			
Maintenance Materials and Supplies			
Labor Overhead	40% of Dir. Lab.		156.1
Misc. Variable Costs/Credits(a)			
Royalty Payments $/yr	108,900		108.9
FIXED COSTS			
Plant Overhead	100% of Dir. Labor/yr		390.3
Local Taxes and Insurance	2% of C.I.		360.0
Depreciation	6.25%/yr		1,125.0
TOTAL PRODUCTION COSTS			3,859.6
Return on Investment (pretax)		20% of C.I.	3,600.0
Subtotal			
POLLUTION CONTROL (incremental)			2,643.0
TOTAL			10,102.6

---

[a] e.g., miscellaneous chemicals, catalysts, supplies, services

e. Technological Considerations

It has been assumed that low-Btu gas could replace refinery gas. Flexigas is autogeneous and Exxon has test-fired low-Btu flexigas in process heaters. However, larger breechings or induced draft fans may be necessary to handle the greater volumes of gas produced. Condensing steam turbine drives were assumed for the gasifier air blowers, which accounts for the large cooling water requirement.

f. Effect on Intermediate and Final Products

As described above, the asphalt feed is converted to low- or high-Btu gas, naphtha, and gas oil. Hence, Flexicoking provides one means for processing asphalt (which is not a fuel) into fuel products. More than 60% of the feed heating value is converted to liquid intermediates -- naphtha and gas oil. Approximately 23% of the feed heating value becomes available as high-Btu gas. Table IV-28 compares the product structure for the base line case with the product structure with Flexicoking of the asphalt included.

5. Process Change B: On-Site Electric Power by Combustion of Vacuum Bottoms

This alternative was evaluated using the Gulf Coast refinery base line.

a. Process Description and Current Status of Technology

In this process option, electric power is generated internally within the refinery rather than purchased from the local power utility. The fuel for this Gulf Coast option is assumed to be asphalt which will require flue gas desulfurization with subsequent sludge disposal.

Figure IV-7 shows a simplified flowsheet of the generating system. Basically the fuel is burned to generate high-pressure steam which is expanded through a conventional steam turbine which provides the motive power for electricity generation.

b. Potential for Energy Conservation

The internal generation of electric power within the refinery does not conserve energy overall -- nor does it consume more energy than when power is purchased, assuming that the internal and external power plants would operate at the same efficiencies. In effect, the form value of the asphalt product is upgraded to a higher form of electric power for refinery use. Table IV-29 shows the net refinery energy balance after implementing option B. Although the ratio of input Btu to output Btu decreases slightly, there is no real net energy cost. This occurs simply because both the energy input decreases (no purchase of electricity) and the output also decreases by the same quantity (asphalt consumed internally).

TABLE IV-28

COMPARISON OF PRODUCT SLATES WITH AND WITHOUT FLEXICOKING
(10^3 BPD)

	East Coast Base Line (1985)	Base Line with Asphalt Flexicoking
LPG	6.2	6.5
High-Btu gas	-	4.3 (FOE)
Unleaded gasoline	106.9	106.9
Naphtha	1.3	1.3
No. 2 distillate	47.0	47.0
Jet fuel	5.9	5.9
Kerosene	3.5	3.5
Low sulfur fuel oil	13.2	13.2
Asphalt	18.7	-
Naphtha intermediate	-	2.1
Gas oil intermediate	-	8.7
Lube stocks	5.1	5.1
BTX	1.4	1.4
Petrochemical feed	4.3	4.3
TOTAL	213.5	210.2

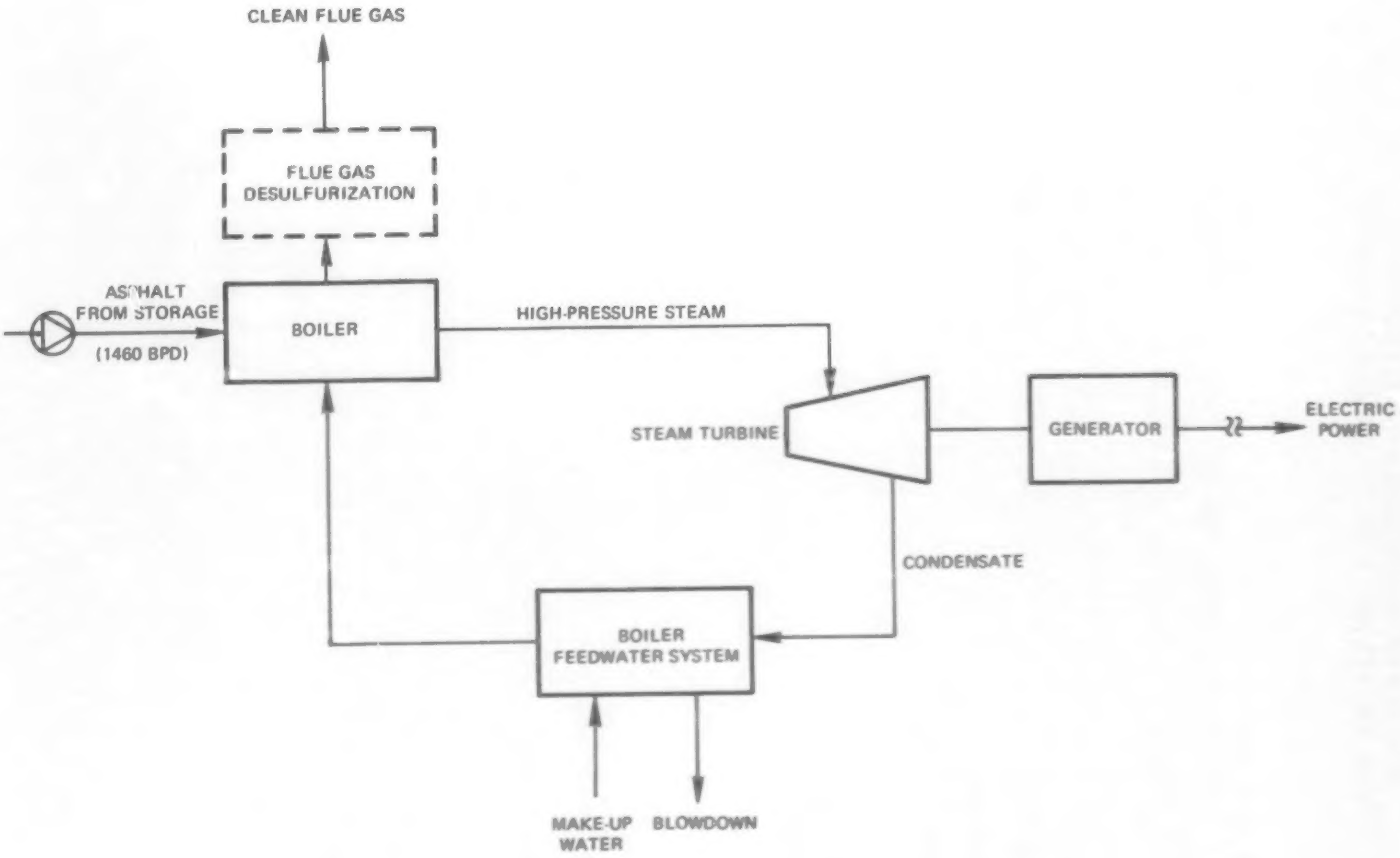

CLEAN FLUE GAS
FLUE GAS DESULFURIZATION
ASPHALT FROM STORAGE (1460 BPD)
BOILER
HIGH-PRESSURE STEAM
STEAM TURBINE
GENERATOR
ELECTRIC POWER
CONDENSATE
BOILER FEEDWATER SYSTEM
MAKE-UP WATER
BLOWDOWN

Figure IV-7. Electric Power from Asphalt

TABLE IV-29

COMPARISON OF REFINERY ENERGY BALANCE WITH AND WITHOUT ON-SITE POWER GENERATION

	Gulf Coast Base Line	Base Line with Onsite Power Generation
Energy In		
Crude	1275.5	1275.5
C_4's	56.1	56.1
Natural gasoline	16.8	16.8
FCC feed (storage)	-	-
Reformer feed	-	-
Electricity	9.2	-
Steam	0.7	0.7
Total	1358.3	1349.1
Energy Out		
Mixed olefins	14.5	14.5
Distillates	1117.8	1117.8
Resid, asphalt, coke	65.5	56.3
Total	1197.8	1188.6
$\frac{\text{(Energy out)}}{\text{(Energy in)}} =$	0.882	0.881

We considered the option of a dual-purpose powerplant in which steam is generated at a high pressure and expanded through a topping turbine for power generation before it is sent to process. The particular ratio of steam-to-electric-power demand for the example refinery, however, would have required that some additional steam be generated and let down through a parallel condensing turbine. In part, for this reason, we chose to consider only an all-condensing power plant.

The other reasons we chose the all-condensing plant relate to the fact that we were considering modification(s) to an existing refinery rather than the grassroots design of a new refinery. Since the existing refinery has in place a steam network of sufficiently high reliability, it would be redundant -- and costly from a capital point of view -- to install new generating capacity for providing steam for the entire refinery instead of that required specifically for electric power generation. Admittedly, this choice precludes the energy savings of the dual-purpose concept, but the capital versus energy cost tradeoffs for an existing refinery will depend on the specific situation which we could not prejudge. Furthermore, vis-a-vis incremental pollution, this option provides a conservative estimate; i.e., more pollution and consequent controls. Nevertheless, further evaluation of integrated energy systems is warranted.

c. Pollution Control

(1) Identification of Pollutants

The pollutants emitted by process change B are indicated on the block flow diagram in Figure IV-8 and described in Table IV-30. Uncontrolled water effluents include cooling water and boiler blowdown. Flue gas from the boiler is the major airborne pollutant.

(2) Water Pollution

The only water pollutional implications of a Gulf Coast refinery employing on-site power generation via asphalt combustion will be an increased cooling tower and boiler blowdown flow rates. The process wastewater flow rate and composition will be essentially the same as the base case. Therefore, the process wastewater treatment cost will be identical to that of the base case.

The cooling tower and boiler blowdown will be increased from 5.19 to 5.71 mgd, and this increase will result in an increase in a total (process plus cooling tower blowdown) wastewater treatment cost of less than 1% (\$4.937 x 10^6/yr versus \$4.895 x 10^6/yr). Cooling tower blowdown treatment costs are shown in Table IV-31.

In terms of both treated effluent waste load and wastewater treatment cost, the implications of implementing on-site power generation via asphalt combustion are negligible.

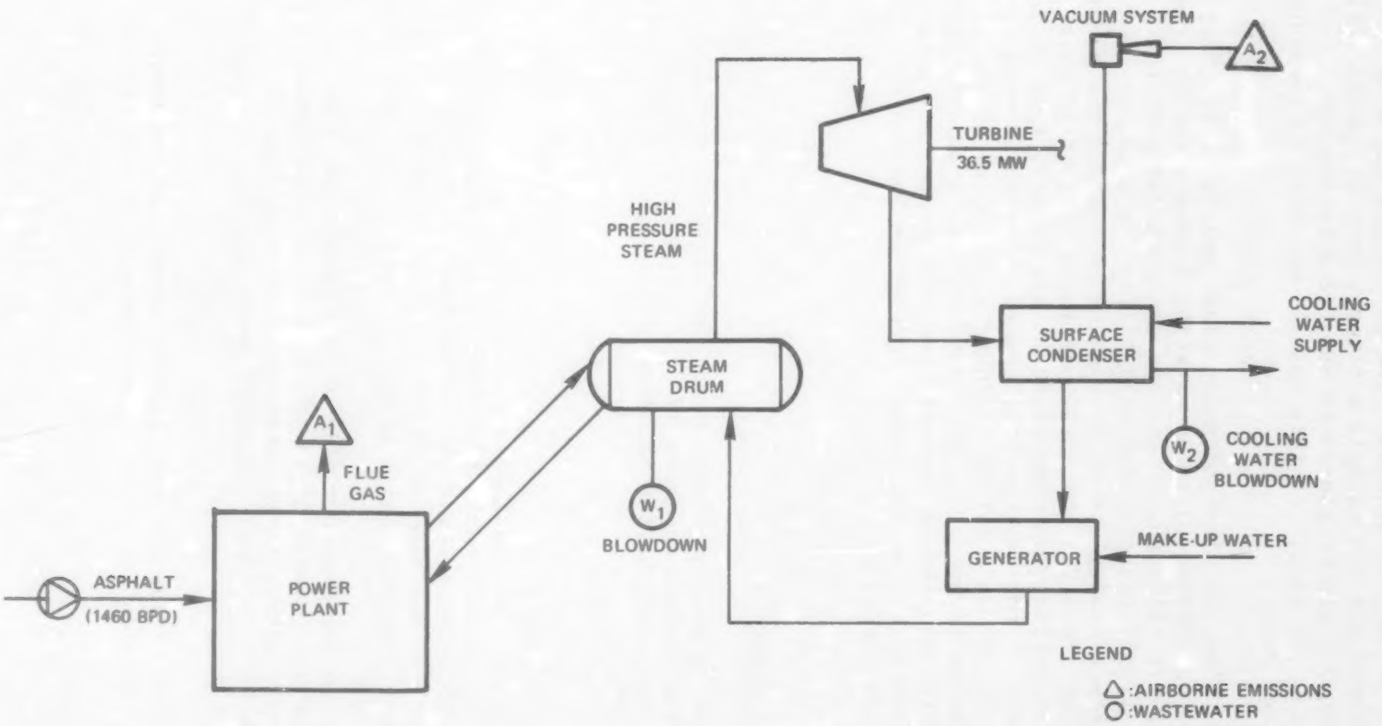

VACUUM SYSTEM
A_2
TURBINE
36.5 MW
HIGH
PRESSURE
STEAM
SURFACE
CONDENSER
COOLING
WATER
SUPPLY
STEAM
DRUM
A_1
FLUE
GAS
W_2
COOLING
WATER
BLOWDOWN
W_1
BLOWDOWN
GENERATOR
MAKE-UP WATER
ASPHALT
(1460 BPD)
POWER
PLANT
LEGEND
△:AIRBORNE EMISSIONS
○:WASTEWATER

Figure IV-8. Pollution Source Identification: Electric Power from Asphalt Combustion

TABLE IV-30

ELECTRIC POWER FROM ASPHALT - POLLUTION PROFILE

Stream No.	Description	Pollutant	Estimated Emission Rate lb/hr Uncontrolled
Water pollution - 10^3 gpd			
W_1	Steam system blowdown	Dissolved solids	192
W_2	Cooling water blowdown	Dissolved solids	333
Air pollution - lb/day			
A_1	Flue gas	SO_x	27,456
		Particulate	1,226
		Hydrocarbons	204
		NO_x	4,234
		Aldehydes	36
A_2	Vacuum off gas (air, steam)	-	-

TABLE IV-31

ALTERNATIVE B

GULF COAST REFINERY WITH ON-SITE POWER GENERATION COOLING TOWER BLOWDOWN WASTEWATER TREATMENT COSTS
(Basis: 200,000 BPD, 330 days/yr)

Capital Investment: $1,390,900

	Yearly Quantity	Cost per Unit Quantity	Yearly cost ($/yr)
Variable Cost			
Operating labor (including supervision overhead)	2200 m-hr/yr	$17.15/m-hr	37,700
Maintenance (including labor and materials)			52,200
Chemicals			
• Sulfur dioxide	473 tpy	$340/ton	160,800
• Lime	505 tpy	$32.50/ton	16,400
• Sulfuric acid	236 tpy	$51.15/ton	12,100
Fuel	-	-	-
Electrical power	119,200 kWh/yr	$0.0145/kWh	1,700
Wastewater treatment sludge disposal	4,660 tpy (@10% solids)	$5.00/ton	23,300
Total variable cost			$304,200
Fixed Cost			
Depreciation @ 6.25%			86,900
Taxes & insurance @ 2%			27,800
Total fixed cost			$114,700
Total annual cost			$418,900
Return on investment @ 20%			$278,200
Total			$697,100

Notes:
1) Capital investment adjusted to the 1975 level (ENR Construction Cost Index - 2126).

2) Wastewater treatment costs are based on a chrome reduction/precipitation system consisting of reaction vessels, clarifiers, chemical feed system and controls.

3) Chromium concentration in untreated cooling tower blowdown is assumed to be 30 mg/l.

4) Cost estimates are for separate treatment of chromium-contaminated cooling tower blowdown. Not all refineries treat cooling tower blowdown separately, and the amount of chromate used varies considerably.

(3) Air Pollution

The major pollutant of concern for option B is the SO_2 emission in the exhaust from the boiler. The cost of sulfur control for on-site electric power generation is shown in Table IV-32. In comparison to the costs shown in Appendix D for flue gas desulfurization of large combustion sources, these costs appear to be somewhat high.

The economy of scale associated with gas desulfurization was discussed with respect to the direct firing of asphalt in process heaters (Option A-1). In the context of that particular option, it appeared to be more economical to sell asphalt as a fuel to large utilities rather than to burn asphalt within the refinery and export fuel oil. Since one might expect the utilities to resist the burning of a difficult fuel such as asphalt, another alternative might be to burn the asphalt at the refinery within a single boiler which would generate both process steam and electricity for the refinery. However, the decision on the part of the refiner to pursue this option, as opposed to purchasing electric power, will be made based upon such factors as prevailing electric rates, reliability of power supplies, and markets for asphalt and coke - and not upon incremental differences for environmental control.

(4) Solid Waste

The solid waste stream for Alternative B - On-site Power Generation - is essentially of the same volume and compositon as that of the base case.

d. Cost Factors

Table IV-33 shows the estimated annual cost of operating the steam/electric system, excluding the cost of fuel. When the cost of pollution control equipment is included, the cost of operation, maintenance, and capital for the power is 1.1¢/kWh not including the fuel charge. This does include a demand charge of $1.86/kW for the Gulf Coast in 1975, which would be paid monthly to the local utility to provide backup electric service in the event of a refinery power failure.

e. Technical Considerations

Since the power generating base of a refinery producing its own power would be much smaller than that of a typical electric utility, the reliability of service would be similarly lower. This is because the utility would normally have several generating stations within its system, each with more than one boiler. The total generating capacity of the utility would exceed the normal expected maximum instantaneous demands and therefore there is a normal built-in reserve. Further, a utility is free to purchase power from the local grid system in the event that demand is unusually high or when several of its generators are out of service.

To assure itself the same general level of reliability of power supply as is available from the utility, the refinery would pay a monthly demand

TABLE IV-32

SULFUR CONTROL COSTS OF ELECTRIC POWER GENERATION (OPTION B)
(Basis: Gulf Coast, 1975 - 383 x 10^6 Btu/hr)

	On-site Electric Generation
Capital cost, $000	
Direct Costs:	
- Scrubber	1,032
- Regeneration[1]	1,484
Total direct costs	2,516
Indirect costs, @ 30% of direct	755
Total capital cost, $000	3,271
Annual operating cost, $000/yr	
Variable costs:	
Labor, 1 man/shift (incl. supv. & overhead) @ $14.85/hr	130
Maintenance, @ 5% of direct capital	126
Utilities	
- Electric power, 240 kWh/10^6 scf @ $0.0145/kWh	127
- Water, 75 gpm @ $0.35/$10^3$ gal	13
- Fuel, 2.5 x 10^6 Btu/ton of S @ $1.73/$10^6$ Btu	10
Total variable costs	406
Fixed costs:	
Depreciation, 16 years	205
Insurance & taxes, @ 2% of capital	65
Total fixed costs	269
Total production cost	675
Return on investment, @ 20% of capital	654
Credit for recovered sulfuric acid, @ $25/ton (net)	(184)
TOTAL ANNUAL COST, $000/yr	1,145
Unit cost, $/$10^6$ Btu	0.38

[1] Based on MgO scrubbing, sulfuric acid recovery.

TABLE IV-33

PRODUCTION COSTS: OPTION B (STEAM/ELECTRIC)

Product: Electric Power
Annual/Design Capacity: 36.5 MW
Process: Steam/Electric (B)
Fixed Investment: $15.8 Million
Stream Days/Yr.: 330
Location: Gulf Coast
Depreciation Period (yr): 16
Year Used for Costing Purposes: 1975

	Units Used in Costing or Annual Cost Basis	$/Unit	Annual Cost ($$10^3$)
VARIABLE COSTS			
Raw Materials			
• Asphalt			
Byproduct Credits			
Energy			
• Purchased Fuel			
• Purchased Steam			
• Electric Power Purchased		$0.03/kWh	
• Miscellaneous			
Energy Credits			
Water			
• Process (Consumption) BFW	70,000 10^3 gal	0.35/10^3 gal	24.5
• Cooling	7.38 × 10^9 gal	$0.044/$10^3$ gal	325.0
Direct Operating Labor (Wages)	35,040 hrs	$6.76/hr	236.9
Direct Supervisory Wages	10% of Dir. Lab.		23.7
Maintenance Labor	1.5% Fixed Inv.		237.0
Maintenance Supervision	10% Main. Lab.		23.7
Maintenance Materials and Supplies	1.5% Fixed Inv.		237.0
Labor Overhead	40% of Dir. Lab.		164.2
Misc. Variable Costs/Credits[a]			
• Demand Charge	36,500 kW	$1.86/kW/mo.	815.0
Royalty Payments			
FIXED COSTS			
Plant Overhead	100% of Dir. Lab.		260.6
Local Taxes and Insurance			316.0
Depreciation			988.0
TOTAL PRODUCTION COSTS			3,591.6
Return on Investment (pretax)			3,160.0
POLLUTION CONTROL (Incremental)			1,189.0
TOTAL			7,940.6

---

[a] e.g., miscellaneous chemicals, catalysts, supplies, services.

charge to the power company which would serve as a rental fee for reserve capacity maintained in the event that it was needed on short notice. This demand charge is included in the economics of Section 5d above.

In addition to paying the demand charge, the refinery would probably hedge its position internally by installing more than one item of smaller equipment in parallel rather than a single, large item to minimize the effect of an outage. Also, some redundancy would be introduced. For example, two boilers of 60-75% capacity each might be installed -- or perhaps three 50% boilers. Because the turbines and generators are typically more reliable than boilers, less overall redundancy would be installed in these areas.

f. Effect on Intermediate and Final Products

Option B has the net effect of reducing the availability for sale of a refinery product (asphalt), while at the same time eliminating the need for normal purchase of a utility (electric power). Otherwise there is no impact on the mix of refinery products.

6. Process Change C: High-Purity Hydrogen Production via Partial Oxidation of Asphalt

This process change was performed for the West Coast refinery cluster.

a. Process Description and Current Status of Technology

This alternative is based on the production of high-purity hydrogen for hydrotreating from vacuum bottoms, using a partial oxidation process such as those offered by Shell and Texaco. The feedstocks freed up by this approach would then be available for sale outside the refinery in the form of pipeline gas or naphtha.

As shown in Figure IV-9, the partial oxidation system consists of several steps. Oil and oxygen are fed to a series of burners in the partial oxidation reactor where, in the presence of steam, a mixture of gaseous hydrogen, carbon monoxide, carbon dioxide, and solid carbon (soot) are produced by the following reactions:

$$C_nH_m + (n/2 + m/4)O_2 \rightarrow nCO + m/2\ H_2O$$

$$C_nH_m + (n + m/4)O_2 \rightarrow nCO_2 + m/2\ H_2O$$

$$CO + H_2O \rightarrow CO_2 + H_2$$

$$2CO \rightarrow CO_2 + C$$

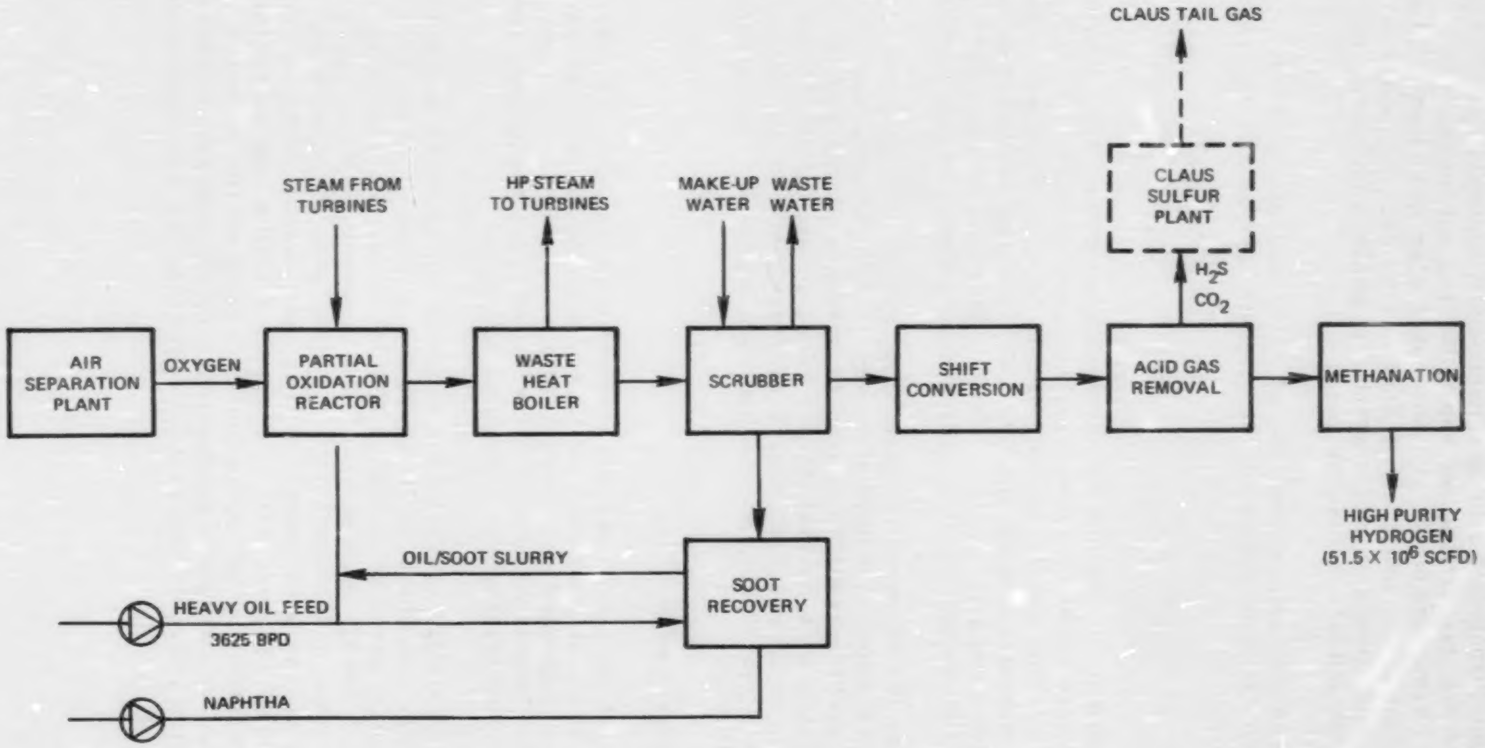

CLAUS TAIL GAS
STEAM FROM TURBINES
HP STEAM TO TURBINES
MAKE-UP WATER
WASTE WATER
CLAUS SULFUR PLANT
H_2S
CO_2
AIR SEPARATION PLANT
OXYGEN
PARTIAL OXIDATION REACTOR
WASTE HEAT BOILER
SCRUBBER
SHIFT CONVERSION
ACID GAS REMOVAL
METHANATION
HIGH PURITY HYDROGEN
(51.5 X 10^6 SCFD)
OIL/SOOT SLURRY
SOOT RECOVERY
HEAVY OIL FEED
3625 BPD
NAPHTHA

Figure IV-9. Hydrogen via Partial Oxidation

The hot reactor effluent gases are cooled by generating high-pressure steam. They then pass through a water scrubber which accomplishes further cooling and saturates the gas with water vapor for the subsequent shift-conversion reaction. Simultaneously, the soot particles are removed from the gas into the liquid phase. By addition of feed oil to the soot/water slurry, the soot leaves the water phase as agglomerates or pellets which may be recycled to the partial oxidation reactor.

The downstream gas processing consists of three additional steps. First, the carbon monoxide is reacted with water to produce hydrogen by the water-gas shift reaction:

$$CO + H_2O \rightarrow CO_2 + H_2.$$

Second, the carbon dioxide is removed by physical absorption, for example, in a hot potassium carbonate solution or chilled methanol (Rectisol). At the same time, H_2S produced from sulfur in the reactor is also absorbed. Upon regeneration, the absorbent releases the CO_2 and H_2S which pass to the existing Claus plant to produce elemental sulfur. Third, the final traces of carbon monoxide are removed by conversion with hydrogen to methane in the methanator.

Partial oxidation (as licensed by Shell and Texaco) has been practiced widely over the past 15 years on feedstocks ranging from natural gas through vacuum bottoms. The systems, usually geared to hydrogen production, have generally proven reliable with availabilities often in excess of 95% for light feedstocks. In short, we do not feel that a refinery would be taking an unwarranted risk in the reliability of its hydrogen supply by installing a partial oxidation unit.

b. Potential for Energy Conservation

The production of hydrogen by partial oxidation of asphalt, as compared with steam reforming of light hydrocarbons (i.e., naphtha or refinery gas), is aimed primarily toward upgrading the overall form values of the refinery output. Converting some of the "bottom of the barrel" for hydrogen production frees up some of the more valuable, lighter fractions for sale.

In terms of actual energy utilization, there is essentially a standoff between the two options. As shown in Table IV-34, a total of about 17×10^6 equivalent Btu/day can be saved by going to partial oxidation. This is less than 0.01% of the total energy equivalent of the energy-related utilities in the refinery, fuel, steam, and electricity. Clearly, this is an even smaller fraction of the total refinery throughput. This very modest energy saving is essentially the result of the fact that, in the base line steam reforming case, CO_2 removal was accomplished by hot potassium carbonate scrubbing which requires process steam for solvent regeneration. In the partial oxidation scheme, the regeneration utilizes the nitrogen byproduct of air separation as the stripping gas.

TABLE IV-34

COMPARISON OF REFINERY ENERGY BALANCE WITH AND WITHOUT HYDROGEN PRODUCTION VIA PARTIAL OXIDATION OF ASPHALT

West Coast - 10^9 Btu/Day

	Base line	Base line with Asphalt POX
Energy In		
Crude	960.6	960.6
C_4's	3.2	3.2
Natural gasoline	5.0	5.0
FCC feed (storage)	20.2	20.2
Reformer feed	10.6	10.6
Electricity	9.5	10.3
Steam	-	(0.1)
Total	1009.1	1009.8
Energy Out		
C_1/C_2	-	23.3
Mixed olefins	7.6	7.6
Distillate products	682.6	682.6
Resid, asphalt, coke	259.6	236.9
Total	949.8	950.4
$\frac{\text{(Energy out)}}{\text{(Energy in)}}$ =	0.941	0.941

c. Pollution Control

(1) Identification of Pollutants

Additional pollutants emitted as a result of adding asphalt partial oxidation to the refinery operations are indicated in Figure IV-10 and Table IV-35. The major pollutant is H_2S contained in the acid gas removed from the raw hydrogen stream.

(2) Water Pollution

It is anticipated that a West Coast refinery employing hydrogen production via partial oxidation will have a treated process wastewater effluent and a treated cooling tower blowdown that will be slightly greater than that of the base case West Coast refinery.

The untreated process wastewater will contain appreciably greater suspended solids; however, these will be easily removed in the existing treatment system, and the suspended solids concentration of the treated effluent will be essentially the same as that of the base case. The increased suspended solids will slightly increase the quantity of wastewater treatment sludge for disposal.

The process wastewater flow rate for the partial oxidation alternative is 5.80 mgd versus 5.78 mgd for the base case and the cooling tower blowdown flow rate is increased from 4.25 mgd to 4.34 mgd. These increased flow rates result in a 2% increase in total wastewater treatment cost over the base case ($\$4.607 \times 10^6$/yr versus $\$4.512 \times 10^6$/yr).

The wastewater treatment cost estimates are presented in Tables IV-36 and IV-37.

In terms of both treated effluent waste loads and wastewater treatment costs, the water pollution implications of implementing the partial oxidation alternative are essentially negligible.

(3) Air Pollution

The major emission associated with this process change is the sulfur removed from the raw syngas in the form of H_2S. This is removed from the gas using an amine scrubber system. The exhaust from the amine regenerator, which contains the sulfur as well as CO_2, is sent to a Claus plant for sulfur recovery and air pollution control. Because several states have emissions standards regulating tailgas for sulfur recovery plants, we have assumed that tailgas cleanup will be required, such that emissions from the plant will be limited to 250 ppm or less. The cost of the sulfur recovery for this option is compared with the base case and the hydrotreating option in Table IV-38. With the partial oxidation case, the sulfur load is increased by only 6%, resulting in a corresponding increase in costs of less than 4%. The major difference between the partial oxidation case and the hydrotreating case is that the partial oxidation process does not use any asphalt as a refinery fuel and, therefore, avoids the high cost of fuel gas desulfurization.

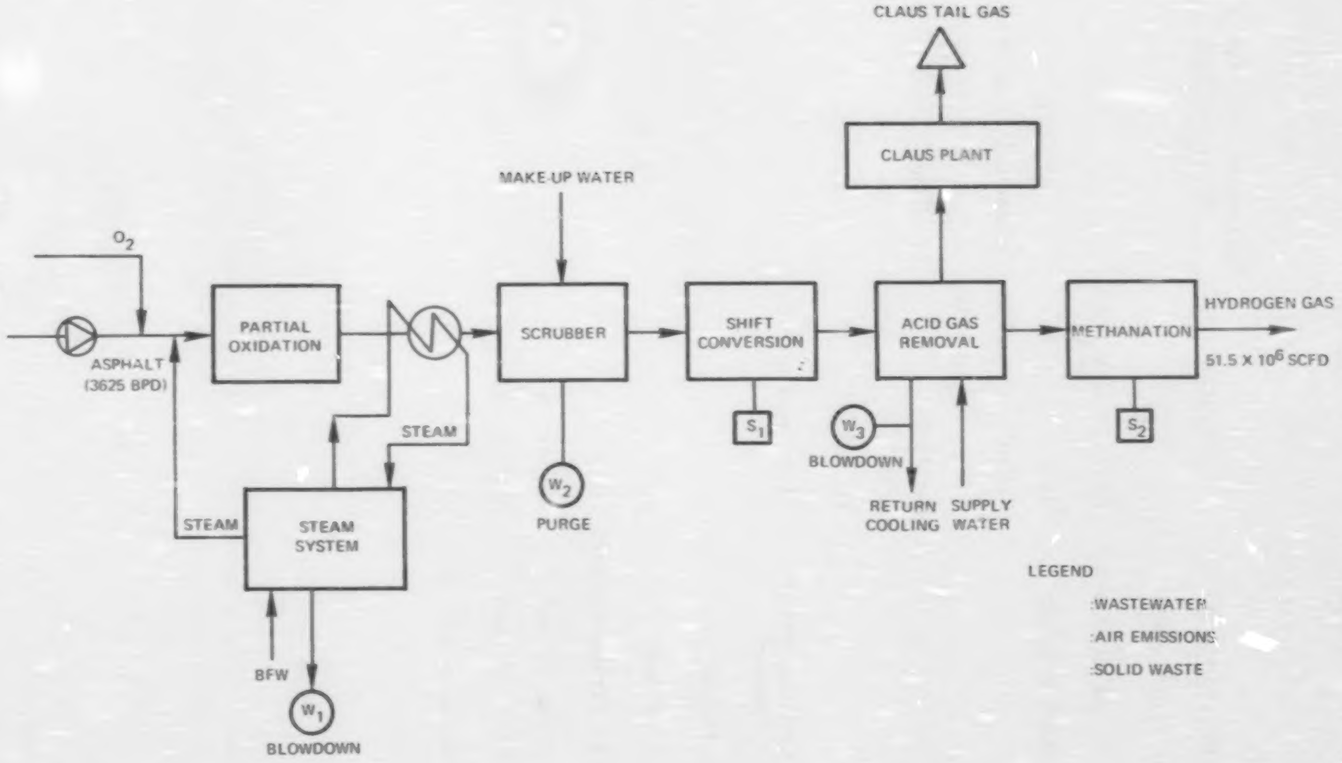

CLAUS TAIL GAS
CLAUS PLANT
MAKE-UP WATER
O_2
ASPHALT
(3625 BPD)
PARTIAL OXIDATION
SCRUBBER
SHIFT CONVERSION
ACID GAS REMOVAL
METHANATION
HYDROGEN GAS
51.5 X 10^6 SCFD
STEAM
STEAM
STEAM SYSTEM
S_1
W_3
BLOWDOWN
RETURN COOLING
SUPPLY WATER
S_2
W_2
PURGE
BFW
W_1
BLOWDOWN
LEGEND
:WASTEWATER
:AIR EMISSIONS
:SOLID WASTE

Figure IV-10. Pollution Source Identification: Partial Oxidation

TABLE IV-35

PARTIAL OXIDATION FOR HYDROGEN PRODUCTION - POLLUTION PROFILE
(Basis: 3625 BPD asphalt)

Stream No.	Description	Pollutant	Estimated Emission Rate Uncontrolled (lb/hr)
Water Pollution - 10^3 gpd			
W_1	Steam blowdown	Dissolved solids	2
W_2	Quench purge	Slurry	22
		Dissolved solids	(leached from ash)
		Suspended solids	3180 lb/day
		Acidity	slight (formic acid)
		HCN	Tr
		Phenolics	Tr
W_3	Cooling water blowdown	Dissolved solids	93
Air Pollution - lb/day			
A_1	Acid gas	H_2S	21,200 (S equiv.)
Solid Waste - tpy (intermittent)			
S_1	HT shift cat.	Fe, Cr, S	36.8
	LT shift cat.	Cu, Zn, S	42.5
S_2	Methanation cat.	Ni	1.5

TABLE IV-36

ALTERNATIVE C

WEST COAST REFINERY - HYDROGEN PRODUCTION BY PARTIAL OXIDATION PROCESS WASTEWATER TREATMENT COSTS

(Basis: 175,000 BPD, 330 days/yr)

Capital Investment: $10,660,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
Variable Cost			
Operating labor (including supervision and overhead)	16,060 m-hr/yr	$17.15/m-hr	275,000
Maintenance (including labor and materials)			426,400
Chemicals			
• Lime	795 tpy	$32.50/ton	25,800
• Phosphoric acid	22 tpy	$440/ton	9,700
• Chlorine	79 tpy	$140/ton	11,100
• Coagulant aid	31,800 lb/yr	$1.00/lb	31,800
• Replacement carbon	108,300 lb/yr	$0.40/lb	43,300
Fuel (gas)	6,500 10^6 Btu/yr	$1.31/$10^6$ Btu	8,500
Electricl power	2,648,500 kWh/yr	$0.0211/kWh	55,900
Wastewater treatment sludge Disposal			
• Incinerator ash	354 tpy	$5.00/ton	1,800
• Stretford purge water	5456 tpy	$15.00/ton	81,800
Total variable cost			$971,100
Fixed Costs			
Depreciation @ 6.25%			666,300
Taxes & insurance @ 2%			213,200
Total fixed cost			$879,500
Total annual cost			$1,850,600
Return on investment			$2,132,000
Total			$3,982,600

Notes: 1) Capital investment adjusted to the 1975 level (ENR Construction Cost Index - 2126).

2) Wastewater treatment costs are based on total implementation (from ground up) of the BATEA level (Best Available Technology Economically Achievable), 1983.

3) Cost estimates are for the specific example, and are in no way intended to represent industry-wide treatment costs.

TABLE IV-37

ALTERNATIVE C

WEST COAST REFINERY WITH PARTIAL OXIDATION HYDROGEN PRODUCTION COOLING TOWER BLOWDOWN WASTEWATER TREATMENT COSTS

(Basis: 175,000 BPD, 330 days/yr)

Capital Investment: $1,308,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
Variable Cost			
Operating labor (including super. and overhead)	2200 m-hr/yr	$17.15/m-hr	37,700
Maintenance (including labor and materials)			52,300
Chemicals			
• Sulfur dioxide	429 tpy	$340/ton	154,900
• Lime	456 tpy	$32.50/ton	14,800
• Sulfuric acid	213 tpy	$51.15/ton	10,900
Fuel	-	-	-
Electrical power	108,200 kWh/yr	$0.0211/kWh	2,300
Wastewater treatment sludge disposal	4.230 tpy (@10% solids)	$5.00/ton	21,200
Total variable cost			$294,100
Fixed Costs			
Depreciation @ 6.25%			81,800
Taxes & insurance @ 2%			26,200
Total fixed cost			$108,000
Total annual cost			$402,100
Return on investment @ 20%			$261,600
Total			$663,700

Notes:
1) Capital investment adjusted to the 1975 level (ENR Construction Cost Index - 2126).

2) Wastewater treatment costs are based on a chrome reduction/precipitation system consisting of reaction vessels, clarifiers, chemical feed system and controls.

3) Chromium concentration in untreated cooling tower blowdown is assumed to be 30 mg/l.

4) Cost estimates are for separate treatment of chromium-contaminated cooling tower blowdown. Not all refineries treat cooling tower blowdown separately, and the amount of chromate used varies considerably.

TABLE IV-38

SULFUR CONTROL COSTS FOR HYDROTREATING (OPTION A-2) AND PARTIAL OXIDATION (OPTION A-3)

	Base line	Hydrotreating	Partial Oxidation
Inlet Sulfur Load, long ton per day	159	182	169
Capital Cost, $000			
Claus plant	3,500	3,850	3,650
Tailgas cleanup	2,500	2,650	2,550
Total	6,000	6,500	6,200
Annual Operating Cost, $000/yr			
Variable costs:			
Labor			
- Direct (1 man/shift @ $6.75/hr)	59	59	59
- Supervision @ 10% of direct labor	6	6	6
Labor overhead @ 40% of labor	26	26	26
Maintenance @ 5% of capital	300	325	310
Utilities			
- Electric power @ 140 kWh/long ton per day, $.0211/kWh	156	178	165
- Fuel @ 0.8 x 10^6 Btu/long ton per day, $2.41/$10^6$ Btu	101	116	108
- Cooling water @ $2.50/long ton per day in tailgas	46	52	49
Total variable costs:	699	762	723
Fixed costs, $000/yr			
Plant overhead @ 100% of labor	65	65	65
Depreciation, 16 years	375	406	388
Insurance & taxes @ 2% of capital	120	130	124
Total fixed costs	560	601	577
Total production cost, $000/yr	1,259	1,363	1,300
Credit for recovered sulfur, @ $25/ton (net)	(1,312)	(1,502)	(1,394)
Return on investment @ 20% of capital	1,200	1,300	1,240
Total annual cost, $000/yr	1,147	1,161	1,146
Estimated incremental cost for flue gas desulfurization @ $0.50/$10^6$ Btu	--	936	--
Total sulfur control cost	1,147	2,097	1,146

(4) Solid Waste

Implementation of hydrogen production via partial oxidation at the base case West Coast refinery will slightly increase the volume of the total solid-waste stream due to the slightly increased quantity of wastewater treatment sludge. In addition, there will be waste catalyst in the amount of approximately 80 tpy.

d. Cost Factors

Table IV-39 shows the net effect on the economics of hydrogen production (excluding feedstock values) caused by shutting down the base line steam reformer and installing a partial oxidation unit. No salvage value is claimed for the reformer which would actually remain on standby for emergency service in the event of a partial oxidation unit outage. Only the savings in actual operating costs are credited to give the net operating cost figures.

e. Technical Considerations

As indicated earlier, the level of overall reliability of partial oxidation of asphalt for hydrogen production will depend largely upon the quality of the feedstock and, at best, will be comparable with steam reforming. As a practical issue, however, we would expect that if a refiner were to shut down his reformer in favor of a partial oxidation unit, he would retain the reformer in a standby mode to effectively increase the overall reliability of hydrogen production. In fact, to some extent these would probably be equipment common to both units. For example, much of the shift conversion, acid gas removal and methanation equipment could be common to both plants. Further, some of the heat recovery equipment could be common. Although in our capital cost estimate we did not account for these factors, they may be valid considerations in a refinery where additional capital investment were to be minimized.

f. Effect on Intermediate and Final Products

Option C is one where, in effect, one refinery product is being substituted for another for the production of an intermediate. The substitution of vacuum bottoms for refinery gas to produce hydrogen is the net effect in this case. The overall refinery energy balance for this option was shown in Table IV-34 in contrast with the base line.

TABLE IV-39

PRODUCTION COSTS: OPTION C (PARTIAL OXIDATION)

Product: Hydrogen

Daily Capacity: 51.47×10^6 scfd

Process: Partial Oxidation

Batch/Continuous:

Fixed Investment: 39.1×10^6

Stream Days/Yr.: 330

Location: West Coast

Depreciation Period (yr): 16

Year Used for Costing Purposes: 1975

	Units Used in Costing or Annual Cost Basis	$/Unit	Annual Cost ($$10^3$)
VARIABLE COSTS			
Raw Materials 3750-3625	(125) FOE/day		
Byproduct Credits			
Energy			
• Purchased Fuel			
• Purchased Steam	(17,200 lb $\times 10^3$)	4.20	72.2
• Electric Power Purchased	29.9×10^6 kWh	0.0211	630.9
• Misc.			
Energy Credits			
Water			
• Process (Consumption)	239×10^6 gal	$0.35/10^3$ gal	83.7
• Cooling (Circulating rate)	2.06×10^9	$0.044/10^3$ gal	90.6
Direct Operating Labor (Wages)	1 man/shift	$6.75/hr	59.1
Direct Supervisory Wages	10% Dir. Op. Lab.		5.9
Maintenance Labor	3% of C.I.		624.0
Maintenance Supervision			
Maintenance Materials and Supplies			
Labor Overhead	40% Dir. Lab.		26.0
Misc. Variable Costs/Credits(a)	24,000		24.0
Royalty Payments			
FIXED COSTS			
Plant Overhead			64.6
Local Taxes and Insurance			782.0
Depreciation 16 yrs. - S.L.			2,444.0
TOTAL PRODUCTION COSTS			4,907.0
Return on Investment (pretax) 20%			7,820.0
POLLUTION CONTROL			110.0
TOTAL			7,930.0

[a] e.g., miscellaneous chemicals, catalysts, supplies, services.

V. IMPACTS OF POTENTIAL INDUSTRY CHANGES

A. IMPACT OF PROCESSING OPTIONS

The five energy conservation alternatives considered in this report focus on internal fuel allocation within the refinery and upgrading of heavy products. In some cases, asphalt is used as a fuel, replacing refinery gas, fuel oil, or purchased power. In other cases, the asphalt is directly converted into a higher fuel form, such as naphtha, gas oil and combustible gas. The products of these conversions are generally available for sale and in one case (Flexicoking) used to replace internal fuel requirements. The impacts these changes have in regard to energy conservation and pollution control costs are assessed in this chapter.

1. Process/Pollution Control Energy Effects

In the previous chapter, frequent reference was made to the fact that none of the options conserves energy in real terms, i.e., the overall thermal efficiency of the refinery is not increased. However, all of the options considered exhibit significant form value conservation by increasing the yields of light petroleum products (refinery gas, distillate products) with a commensurate reduction in heavy-ends (asphalt, residual oil). The net result of these changes on the refinery product slates is summarized in Table V-1.

The large increase in refinery gas available from option A-3 is a result of displacing refinery gas with low calorific value flexigas and the simultaneous production of high-Btu coker gas. The gas available is equivalent to 28 million scf/day of pipeline gas, which could be sold locally, over the fence, to a gas-dependent industrial consumer. This holds for options A-1 and C as well. Options B and C essentially trade asphalt for another energy form, viz., electricity and refinery gas.

The effectiveness of these changes on the overall refinery energy balance is generally slight, as indicated in Table V-2. As one might expect, the H-oil and Flexicoking conversion processes have the most impact upon energy consumption. In these two cases, there is a net loss in available energy, as contrasted with the other options, where only the form of the energy consumed changes. The ratios presented in Table V-2 are exclusive of energy consumed by pollution control systems.

For all of the options assessed, we concluded that the control of wastewater effluents has a negligible impact. As indicated in Appendix D, the incremental wastewater loads associated with these process changes are small

TABLE V-1

CHANGES TO PRODUCT SLATES RESULTING FROM PROCESS CHANGES
(BPD)

OPTIONS:	A-1 Direct Combustion		A-2 H-Oil		A-3 Flexicoking		B Power Generation		C Asphalt POX	
	Reduction	Increase	Reduction	Increase	Reduction	Increase	Reduction	Increase	Reduction	Increase
Products										
Asphalt (vac. btms.)	12,330	---	9,240	---	18,700	---	1,460	---	3,625	---
Tar	---	---	---	1,480	---	---	---	---	---	---
Residual oil (atm. btms.)	---	9,640	2,070	---	---	---	---	---	---	---
Heavy gas oil	---		---	3,820	---	8,746	---	---	---	---
Light gas oil	---		---	3,060	---	---	---	---	---	---
Naphtha	---	---	---	2,270	---	2,111	---	---	---	---
LPG (FOE)	---	---	---	540	---	350	---	---	---	---
Refinery gas (FOE)	---	2,690	800	---	---	4,489[1]	---	---	---	3,750
Coke	---	---	---	---	---	374	---	---	---	---
Electricity	---	---	---	---	---	---	---	1,460[2]	---	---
Total	12,330	12,330	12,110	11,170	18,700	16,070	1,460	1,460	3,625	3,750

---

[1] Includes displaced refinery gas and coker product gas.

[2] 1,460 $(6.3 \times 10^6/0.0105 \times 10^6)$ = 876,000 kWh/day = 36.5 MW of capacity.

relative to the base line cases. Consequently, there is an insignificant effect on the characteristics of the treated effluents, the type of treatment required to conform with regulatory requirements and the incremental energy consumed.

None of the process changes introduces new and different substances into the wastewater, and none of the process changes significantly alters the volumes of wastewater that must be treated. In essence, we anticipate that the nature of the treated effluent will be very nearly the same as that delineated for the base case.

Conversely, the control of airborne emissions, in this case SO_x, has an identifiable impact in terms of both energy consumption and cost. A summary of the energy consumed by the air pollution control (APC) systems is presented in Table V-3 and compared to the energy (utilities/fuel) consumed by the base line refineries. Except for option A-1, the APC system energy consumption is a very small percentage of the total refinery utility and fuel energy consumption. Option A-1 requires the installation of several small-scale flue gas desulfurization (FGD) systems on process heaters and boilers which results in a relatively higher energy consumption than expected for the other options.

2. Cost Impact

The net cost of pollution controls for each option based upon the cost estimates of Chapter IV is presented in Table V-4. The differential cost for controlling water, air, and solid-waste pollution is developed for each case. The total cost of pollution control is then compared to the operating costs (less feedstock and product values) associated with the process option. The cost impact of pollution controls on option A-1 is substantial, amounting to over 50% of the total operating cost of this alternative. This result is perhaps indicative of why refiners prefer to control refinery sulfur emission by blending high/low sulfur fuels rather than retrofitting FGD systems and burning high sulfur oil.

Pollution controls add about 25 percent to the costs for Flexicoking. In this case, about 30% of the pollution differential is for wastewater treatment. This is mainly associated with increased cooling tower blowdown resulting from the use of condensing steam turbine drives on the air blowers. The remainder is primarily the operating cost for the Stretford system used to treat the flexigas before combustion. The pollution costs associated with option C are small since the increased amount of acid gas generated represented only a small increment on the existing Claus plant.

It should be recalled that these options were analyzed within the context of geographically and functionally different refinery clusters, so that the costs are only broadly comparable with each other.

B. SYSTEMS IMPLICATIONS

In broad terms, the objective of these process changes is to obtain more usable clean energy out of a barrel of crude oil. In some cases, the implementation of these changes has implications beyond the confines of the refinery.

TABLE V-2

CHANGE IN REFINERY ENERGY USAGE RESULTING FROM PROCESS CHANGES

		Refinery Energy Ratio[1]	
Option	Location	Base line[2]	Base line with Process Change
A-1	East Coast	0.904	0.903
A-2	West Coast	0.941	0.935
A-3	East Coast	0.904	0.883
B	Gulf Coast	0.882	0.881
C	West Coast	0.941	0.941

[1] Heating value of products/feed plus utilities.

[2] Refinery fuel in compliance with state regulations, 99.5% overall sulfur recovery from acid gas streams and no SO_x control FCC catalyst regenerator.

TABLE V-3

ENERGY CONSUMED BY AIR POLLUTION CONTROL (APC) SYSTEMS

10^9 Btu/day

Options	A-1	A-2	A-3	B	C
Electricity[1]	2.60	0.04	0.70	0.31	0.02
Fuel	0.13	0.02	0.25	0.02	0.01
Total	2.73	0.06	0.95	0.33	0.03
Baseline utilities and fuel	116.1	127.1	116.1	118.7	127.1
APC energy, % of base line	2.3	0.05	0.8	0.3	0.02

---

[1] 10,500 Btu/kWh used to convert to heat equivalency.

TABLE V-4

SUMMARY OF ANNUAL POLLUTION CONTROL COSTS

($000)

Option	A-1	A-2	A-3	B	C
Water Pollution					
With option change	5032	4613	5735	4963	4647
Base line	(5032)	(4541)	(5032)	(4919)	(4541)
Differential	0	74	703	44	106
Air Pollution					
With option change	12,048[1]	2097	2840	1145	1145
Base line	9,997[2]	(1147)	(1130)	-	(1145)
Differential	2,051	950	1710	1145	0
Solid Waste					
Differential	-	2	230	-	4
Total differential	2051	1026	2643	1189	110
Cost of pollution control, $000/day	6.2	3.1	8.0	3.6	0.3
Operating costs for options, $000/day	5.6	19.3	22.6	20.4	38.6
Total PC & operating cost, $000/day	11.8	22.4	30.6	24.0	38.9
Control cost, % of total	53	14	26	15	<1

---

[1] Assumes pollution control by FGD @ $0.47/10^6 Btu.
[2] Assumes pollution control by HDS and blending @ $0.39/10^6 Btu.

1. Direct Combustion of Asphalt

The burning of heavy, high-sulfur residues in the refinery would increase the availability of high-Btu gas and low-sulfur oil to products. The consequence of this approach is to place the cost of pollution control on the refiners, who are by virtue of available technology in the best position to deal with it. However, the refiner must have an incentive to make this change, viz., the prospect of recovering the additional cost in the sale of products; consequently, the end-user must expect to pay a higher price for fuel. Otherwise the refiner will stay with the present approach of blending fuels.

Another alternative is to sell asphalt to a large energy user such as an electric utility. The burning of asphalt in large boilers with pollution controls would thereby achieve lower unit costs as a result of the improved economies of scale. However, this presents a potential conflict with the nation's implied energy policy to require utilities to burn coal instead of oil.

2. On-Site Power Generation

By choosing this option, the refinery is competing with the large integrated system of the electric utility. The implications of this high-sulfur fuel burning option are basically similar to those of direct combustion discussed above, with the exception of the integrated steam/electrical system.

3. Upgrading of Heavy Residues (H-oil, Flexicoking, Asphalt POX)

The options included in this category appear to be the most practical in terms of implementation and pollution impact. Because these options convert low-quality petroleum residues to higher quality and value intermediate and final products, a refiner has an incentive to apply these alternatives, providing there is sufficient product demand. In addition, the pollution control requirements for these options are not unreasonably expensive and, in general, represent only small incremental changes to already existing facilities.

C. FACTORS AFFECTING PROBABILITY OF CHANGE

1. Direct Combustion of Asphalt

It was clear from our interviews with industry representatives that refiners place a relatively high value on refinery gas, and selling it for outside consumption is an alternative which generates little enthusiasm. There are two possible situations which could bring about a change in this attitude.

- Increased gas prices allowed by deregulation, due to severe shortages; and
- End-use controls on methane gas.

While such occurrences are possible, it appears that the cost of FGD would, in most cases, discourage direct combustion of asphalt as a means of increasing gas availability.

2. Power Generation

The most important factor affecting the implementation of this option would be the setting of utility rates for industrial uses. Recently there has been a lot of public consternation regarding the present inverted rate structure which favors large industrial users. Pressure is building, particularly on the East and West Coasts, to adopt a uniform rate structure. Depending on the outcome, this option could be selected by refiners.

Again, the cost of FGD has an identifiable impact on combustion of asphalt for this purpose. Consequently, internal generation of electricity would most likely be implemented by integrating a topping cycle into the refinery high-pressure steam/waste heat recovery system, since the refinery has many uses for process steam.

3. Upgrading of Asphalt

The factors affecting the choice of upgrading processes for heavy residues are many and complex (as indicated in section IV-B). Important factors include the quality of the crudes processed by the refinery and the demands for the various products, recognizing that the product yields from the available processes also vary widely. With crude quality in U.S. refineries generally on the decline, and with unimpressive growth prospects for asphalt and coke, we expect that more heavy residue conversion will be practiced.

D. AREAS FOR RESEARCH

In the petroleum refining sector, there are two main technology research areas which could produce benefits for the nation in terms of energy availability and improved environmental quality. The first is improving the reliability and reducing the cost of flue gas desulfurization systems, especially in the size range below 50 MW equivalent. The second is to develop rugged hydrocracking catalysts capable of withstanding the poisonous metals which concentrate in petroleum residues or, alternatively, developing a low-cost residual demetallizing process which would, in effect, have the same result, viz., increased catalyst activity and product yields. Technology improvements in either area would increase the yield of clean fuels by permitting greater flexibility in the use of petroleum heavy-ends, and simultaneously ensuring environmental quality.

REFERENCES

Bryant, H. S., "Environment Needs Guide Refinery Sulfur Recovery," The Oil & Gas Journal, March 1973.

Burchard, J. K., Some General Economics Considerations of Flue Gas Scrubbing for Utilities, Control Systems Division, EPA.

U.S. Environmental Protection Agency, Compilation of Air Pollution Emission Factors, AP-42 EPA, 3nd ed., April 1973.

U.S. Environmental Protection Agency, Development Document for Proposed Effluent Limitations Guidelines and New Source Performance Standards for Petroleum Refining, Effluent Guidelines Div., EPA. 440/1-73/014.

Goar, B. G., "Cost, Air Regulations Affect Process Choice," The Oil & Gas Journal, August 1975.

Arthur D. Little, Inc., The Impact of Lead Additive Regulations on the Petroleum Refining Industry, EPA Contract 68-02-1332, Task Order No. 7, Final Report, 1975. (ADL, 1975a)

Arthur D. Little, Inc., The Impact of Producing Low-Sulfur, Unleaded Motor Gasoline on the Petroleum Refining Industry, Contract 68-02-1332, Task Order No. 8, Final Report, 1975. (ADL, 1975b)

Effluent Guidelines and Standards - Petroleum Refining, 40 CFR 419 Fed. Reg., May 1974.

APPENDIX A

INDUSTRY STRUCTURE

1. DESCRIPTION

The petroleum refining industry is perhaps one of the most complex and technically sophisticated industries in the United States. There are some 250 to 300 refineries in the United States ranging in size from about 400,000 BPD to only a few hundred BPD. These refineries vary from a fully integrated, high-complexity plant capable of producing a complete range of petroleum products and some petrochemicals to very simple plants capable of producing only a very small number of products. Also some refineries are modern and of recent construction, while others contain at least some operating process units constructed 40 or more years ago. A survey of operating refineries in the United States between 1962 and 1972 is presented in Table A-1. The crude slates for refineries vary widely and the product mixes and, to some extent, product properties also vary from refinery to refinery. Because of this, each refinery is characterized by a unique capacity, processing configuration, and product distribution.

For the purpose of collecting statistics on the refining industry, the United States has been divided into several refining regions called Petroleum Administration for Defense (PAD) districts. Figure A-1 indicates which states are included in each region. A breakdown of refining capacities and actual crude runs by PAD districts for the period 1960 through 1974 are shown in Table A-2. The major growth in capacity over this period has been in PAD's II, III, and V, as one might expect.

It is apparent that the refining industry is characterized by a diversity of operations; however, within a given PAD district, there are similarities.

ADL has developed a computer simulation of the refining industry based upon six cluster models which represent geographic regions (PAD's) in the United States having refineries with similar characteristics. These six models have been successfully used to simulate the 250 to 300 individual refineries and hence the U.S. refinery industry (EPA 68-02-1332 Task Order No. 7). Generally each cluster model corresponds to a PAD district; however, PAD III, representing 40% of the U.S. total refining capacity, was simulated by two models, while PAD IV (Rocky Mountains) represented such a small percent of the refining industry (less than 5% of the total) that it was not simulated at all.

TABLE A-1

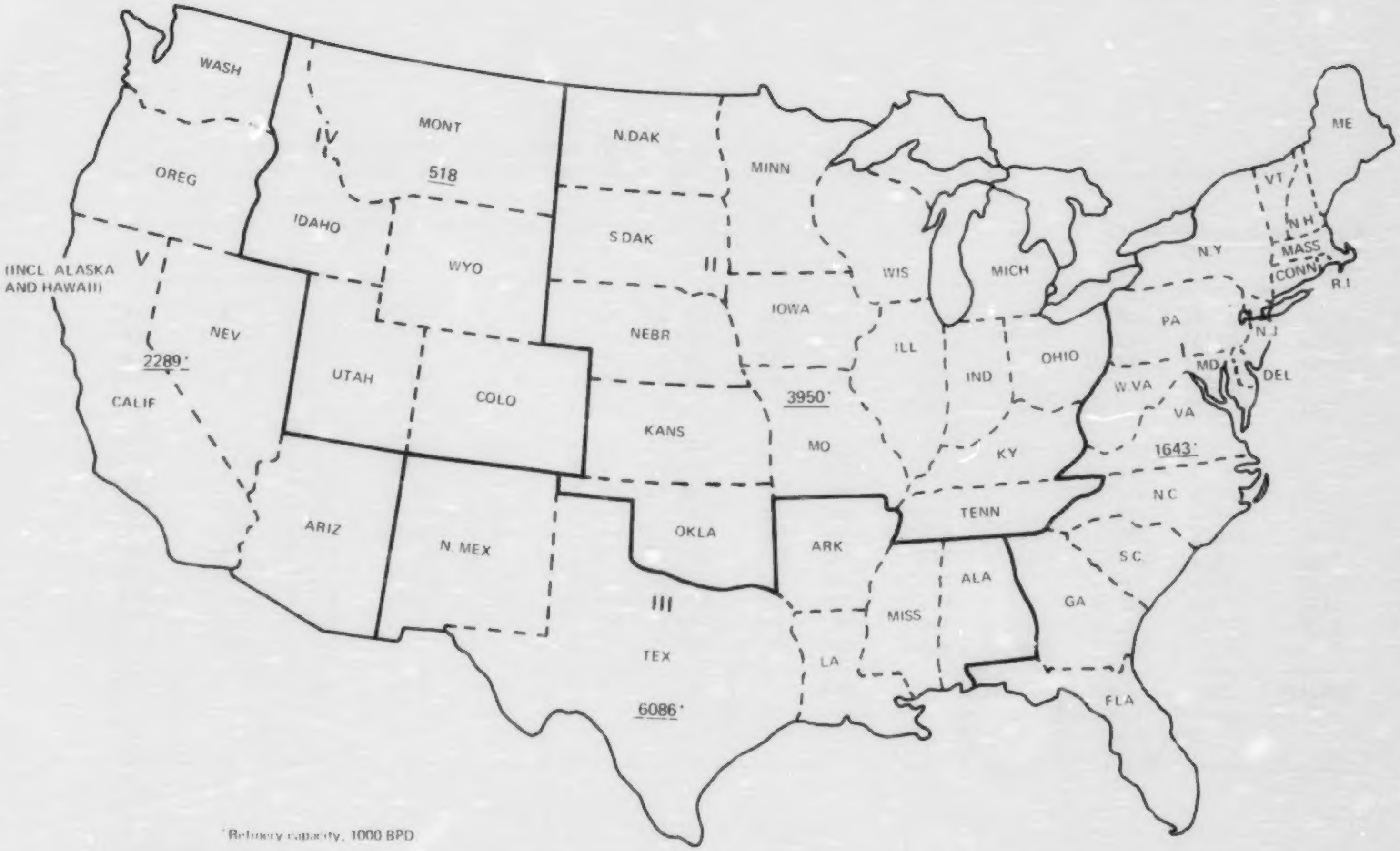
SURVEY OF OPERATING REFINERIES IN THE UNITED STATES - 1962-1972

				Charge Capacity (BPSD x 10^6)								Production Capacity (BPSD x 10^6)				
		Operating Refining Capacity				Catalytic Cracking										
Date	Number Plants	BPCD x 10^6	BPSD x 10^6	Vacuum Distillation	Thermal Operation	Fresh Feed	Recycle	Catalytic Reforming	Catalytic Hydro-cracking	Catalytic Hydro-refining	Catalytic Hydro-treating	Alkyla-tion	Catalytic Polymeriza-tion*	Lube	Asphalt	Coke (MT/SD)
1/1/62	299	10.01	10.59	3.67	1.81	3.75	1.47	2.02	-	-	2.37	0.46	0.14	0.21	0.49	18.90
1/1/63	293	9.92	10.46	3.58	1.75	3.89	1.55	1.99	-	-	2.54	0.49	0.13	0.20	0.49	19.20
1/1/64	288	10.18	10.72	3.75	1.72	3.99	1.62	2.05	-	-	2.75	0.50	0.13	0.20	0.51	20.94
1/1/65	275	10.25	10.76	3.76	1.64	3.99	1.57	2.06	-	-	2.93	0.53	0.13	0.21	0.54	21.14
1/1/66	265	10.25	10.75	3.76	1.69	3.96	1.53	2.09	-	-	3.10	0.55	0.12	0.21	0.53	23.03
1/1/67	261	10.45	10.95	3.89	1.64	3.95	1.65	2.19	-	-	3.35	0.60	0.11	0.21	0.54	25.00
1/1/68	269	11.14	11.66	4.08	1.66	4.18	1.60	2.38	0.41	-	3.66	0.65	0.10	0.21	0.53	28.43
1/1/69	263	11.57	12.08	4.12	1.60	4.25	1.55	2.54	0.50	0.55	3.27	0.67	+0.25	0.20	0.57	29.43
1/1/70	262	12.15	12.65	4.55	1.64	4.37	1.49	2.78	0.60	0.54	3.51	0.75	0.29	0.21	0.58	35.49
1/1/71	253	12.68	13.28	4.74	1.56	4.51	1.46	2.89	0.73	0.54	3.81	0.78	0.31	0.22	0.60	38.77
1/1/72	247	13.09	13.71	4.85	1.53	4.57	1.26	3.17	0.84	0.63	4.26	0.82	0.29	0.22	0.62	41.47
Incremental Change																
1962-1972		3.08	3.12	1.18	(0.28)	0.82	(0.21)	1.15	0.43	0.08	1.89	0.36	-	0.01	0.13	22.57
% Crude		100.0	101.3	37.8	(9.0)	26.3	(6.7)	36.9	13.8	2.6	60.6	11.5	-	0.3	4.2	-
NPC Questionnaire & Expansion Data (MMB/CD)																
1973-1978		1.8	-	0.50	(0.18)	0.20	0.0	0.67	0.11	0.67	0.99	0.13	0.14	0.18	0.09	6.4
% Crude		100.0	-	28.0	(1.0)	11.0	0.0	37.0	6.0	37.0	55.0	7.0	8.0	1.0	5.0	-

---

*Aromatic and isomerization reported beginning 1/1/69.

Source: Oil and Gas Journal, Annual Refining Reports (January 1, 1962 through January 1, 1972); NPC Refining Survey Questionnaire (1973 through 1978).


WASH
OREG
IV
MONT
518
IDAHO
WYO
(INCL. ALASKA AND HAWAII)
V
NEV
2289*
CALIF
UTAH
COLO
ARIZ
N. MEX
N.DAK
S.DAK
NEBR
KANS
OKLA
III
TEX
6086*
MINN
II
IOWA
3950*
MO
ARK
LA
WIS
ILL
MICH
IND
OHIO
KY
TENN
MISS
ALA
GA
FLA
S.C.
N.C.
1643*
VA
W.VA
MD
DEL
N.J.
PA
N.Y.
CONN
R.I.
MASS
N.H.
VT
ME

*Refinery capacity, 1000 BPD

Source: Bureau of Mines

Figure A-1. Petroleum Administration for Defense (PAD) Districts

TABLE A-2

U.S. PETROLEUM REFINING CAPACITY AND ACTUAL CRUDE RUNS AVERAGE JANUARY 1 CAPACITY IN GIVEN YEAR AND JANUARY 1 CAPACITY IN THE FOLLOWING YEAR

(BPCD x 10^3)

Year	District I		District II		District III		District IV		District V		Total U.S.		
	CAP	RUNS	CAP	RUNS	CAP	RUNS	CAP	RUNS	CAP	RUNS	CAP	RUNS	PERCENT
1960	1,537	1,217	2,766	2,415	3,464	2,983	330	285	1,488	1,166	9,585	8,066	84.2
1961	1,558	1,223	2,784	2,433	3,520	3,026	345	284	1,515	1,219	9,722	8,185	84.2
1962	1,543	1,214	2,809	2,463	3,584	3,178	360	304	1,519	1,252	9,815	8,411	85.7
1963	1,493	1,246	2,847	2,541	3,671	3,325	375	307	1,555	1,268	9,941	8,687	87.4
1964	1,464	1,212	2,870	2,570	3,805	3,410	385	319	1,589	1,295	10,113	8,806	87.1
1965	1,420	1,200	2,896	2,645	3,864	3,506	386	330	1,600	1,362	10,166	9,043	89.0
1966	1,400	1,259	2,951	2,790	3,906	3,665	386	338	1,649	1,392	10,292	9,444	91.8
1967	1,416	1,267	3,049	2,849	4,228	3,903	394	344	1,707	1,452	10,794	9,815	90.9
1968	1,452	1,303	3,161	2,959	4,581	4,118	413	366	1,770	1,567	11,377	10,313	90.6
1969	1,477	1,309	3,220	3,013	4,737	4,264	422	385	1,875	1,658	11,731	10,629	90.6
1970	1,487	1,290	3,360	3,154	5,027	4,357	424	393	1,974	1,676	12,272	10,870	88.6
1971	1,515	1,330	3,482	3,230	5,345	4,492	424	405	2,081	1,742	12,847	11,199	87.2
1972	1,566	1,318	3,567	3,311	5,498	4,817	438	397	2,178	1,853	13,247	11,696	88.3
1973	1,638	1,501	3,777	3,485	5,733	5,054	479	415	2,212	1,975	13,839	12,430	89.8
1974	1,643	1,409	3,950	3,333	6,086	5,103	518	421	2,289	1,869	14,486	12,135	83.8

CAP -- Capacity of operating refineries. Source: U.S. Bureau of Mines. Data for 1975 and later obtained by adding known projects from Table A-11.

RUNS -- Crude runs for year indicated from U.S. Bureau of Mines...1975 per Office of Oil and Gas estimate. 1975 and later at 90 percent utilization is illustrative and not a forecast.

Source: FEA

The major factor used by ADL in selecting the refineries in a cluster was the process configuration. Hence these clusters are indicative of the types of refineries in the different PAD regions of the United States today, except for small (less than 30,000 BPD) atypical plants. These are discussed later.

In addition to the typical refineries, there are a large number of small, non-integrated, essentially unique refineries in the United States. These small refineries represent well over one half of the total number of those in the United States. However, their combined process capacity is less than 10% of the total U.S. throughput. The plants range from those whose "primary" product is asphalt to those that are essentially topping refineries producing a very small number of products. Often these refining configurations are designed to enable them to produce products for marketing in the immediate geographic area. These can be viewed as specialty refineries. These refineries are essentially incapable of simulation or easy description, except on a case-by-case basis. Their energy consumption per barrel of feed or product will vary widely as will their pollution emissions.

2. ECONOMIC OUTLOOK

Historically, the number of refineries in the United States has decreased over the last few decades, and the average size has increased dramatically during the same time period as indicated in Table A-3. During the past 4-5 years or so, there have been few new grass root refineries built in the United States. There have been several major expansions to existing refineries and, of course, many revamps of existing units to reflect changing technology. There are currently no new refineries under construction in the United States, although there are several in the planning stage, including at least one in the northeast (Table A-4). There have also been several refinery projects, both grass roots and modernization, cancelled or greatly delayed in the last few years (Table A-5), because of environmental constraints. Many of these were in the northeast where environmental concern is great.

The companies interviewed by ADL did not anticipate rapid growth in the U.S. refining capacity between now and 1985. Some companies interviewed even expected no refinery growth (in crude capacity) before 1985 within the United States. They expected the growth to occur outside the United States. There would be downstream (of the crude unit) additions, alterations, and the like. One reason for the anticipated limited growth in capacity was forecast of nearly zero growth in gasoline demand (see later). A second reason was the rapid increase in cost of a modern refinery - from \$2000/bbl of daily capacity a few years ago to \$3000 to \$4000/bbl now, i.e., a 100,000-BPD refinery, perhaps the smallest practical size, would cost \$400,000,000. Each

TABLE A-3

HISTORICAL REFINING CAPACITY DATA - TOTAL UNITED STATES

	No. of Operating Refineries	Beginning Operating Capacity	Shutdown Capacity		Additions			Grassroots			Total Additions		
		BPD	BPD	%	No.	BPD	%	No.	BPD	%	No.	BPD	%
1962	287	9,793,748*	292,899	3.0	64	211,119	2.2	4	26,500	.3	68	237,619	2.4
1963	287	9,814,791	303,530	3.1	61	233,775	2.4	3	142,100	1.4	64	375,875	3.8
1964	282	10,063,164	322,210	3.2	54	357,731	3.6	6	197,950	2.0	60	555,681	5.6
1965	273	10,161,311	613,284	6.0	56	139,596	1.4	1	700	.0	57	140,296	1.4
1966	267	10,171,159	321,580	3.2	82	428,663	4.2	1	1,000	.0	83	429,663	4.2
1967	260	10,412,447	347,160	3.3	91	665,313	6.4	12	202,250	1.9	103	867,563	8.3
1968	270	11,172,694	360,160	3.2	88	492,114	4.4	1	800	.0	89	492,914	4.4
1969	264	11,575,829	163,680	1.4	73	412,146	3.5	1	89,000	.8	74	501,146	4.3
1970	262	11,882,393	191,930	1.6	81	814,149	6.9	3	280,000	2.4	84	1,094,149	9.2
1971	253	12,658,248	361,730	2.9	54	418,250	3.3	5	272,500	2.2	59	690,750	5.5
10-Yr Summary**		9,793,748			704	4,172,856	42.7	38	1,212,800	12.4	742	5,385,656	55.0
Compounded Annual Rate of Increase or (Decrease)							3.6			1.2			4.5

	Capacity Lost Through Total or Partial Refinery Shutdowns			Capacity Lost Through Consolidations			Other Declines in Capacity			Total Declines			Net Increase or (Decrease)		Ending Operating Capacity
	No.	BPD	%	No.	BPD	%	No.	BPD	%	No.	BPD	%	BPD	%	BPD
1962	5	57,300	.6	3	10,250	.1	31	149,026	1.5	39	216,576	2.2	21,043	.2	9,814,791
1963	8	49,900	.5	1	1,100	.0	21	76,502	.8	30	127,502	1.3	248,373	2.5	10,063,164
1964	14	113,044	1.1	9	273,280	2.7	28	71,210	.7	51	457,534	4.5	98,147	1.0	10,161,311
1965	4	44,300	.4	2	7,000	.1	19	79,148	.8	25	130,448	1.3	9,848	.1	10,171,159
1966	6	57,200	.6	2	40,500	.4	21	90,675	.9	29	188,375	1.9	241,288	2.4	10,412,447
1967	2	3,500	.0	2	80,400	.8	18	23,416	.2	22	107,316	1.0	760,247	7.3	11,172,694
1968	10	46,650	.3	1	2,000	.0	19	41,129	.4	30	89,779	.7	403,135	3.6	11,575,829
1969	6	40,600	.4	1	20,600	.2	22	133,382	1.2	29	194,582	1.7	306,564	2.7	11,882,393
1970	11	102,300	.9	2	120,900	.1	18	95,094	.8	31	318,294	2.7	775,855	6.5	12,658,248
1971	7	58,300	.5	1	72,000	.6	29	183,880	1.4	37	314,180	2.5	376,570	3.0	13,034,818
10-Yr Summary**	73	573,094	5.9	24	628,030	6.4	226	943,462	9.6	323	2,144,586	21.8	3,241,070	33.1	13,034,818
Compounded Annual Rate of Increase or (Decrease)			.5			.6			.9			2.0		2.9	

*1/1/62 capacity later revised by Bureau of Mines to 9,812,248 BPD.

**10-year summary percentages are based on 1962 beginning operating capacity.

Source: U.S. Bureau of Mines, "Petroleum Refineries in the United States", Mineral Industries Survey, published annually.
NPC

TABLE A-4

NEW REFINERIES, EXPANSIONS AND REACTIVATIONS SCHEDULED IN THE UNITED STATES BY PAD DISTRICTS

(BPCD x 10^3 of Crude Distillation)

Type Code*	Company/Location	PAD Districts					Total
		I	II	III	IV	V	
	1975						
E	Socal (Perth Amboy, NJ)	80.0	--	--	--	--	80.0
E	Standard of Indiana (Baltimore, MD)	4.0	--	--	--	--	4.0
E	Clark (Hartford, IL)	--	22.0	--	--	--	22.0
E	CRA (Phillipsburg, KS)	--	4.5	--	--	--	4.5
E	Bay Refining (Bay City, MI)	--	10.0	--	--	--	10.0
E	Conoco (Ponca City, OK)	--	7.5	--	--	--	7.5
E	Kerr McGee (Wynnewood, OK)	--	15.1	--	--	--	15.1
E	Gladieux (Fort Wayne, IN)	--	3.5	--	--	--	3.5
E	Vickers (Ardmore, OK)	--	32.5	--	--	--	32.5
E	United Refining (West Branch, MI)	--	8.5	--	--	--	8.5
E	Lakeside Refining (Kalamazoo, MI)	--	1.6	--	--	--	1.6
R	Saber (Corpus Christi, TX)	--	--	10.5	--	--	10.5
N	Sigmor (Three Rivers, TX)	--	--	5.0	--	--	5.0
E	Good Hope (Good Hope, LA)	--	--	20.1	--	--	20.1
E	Marion (Theodore, AL)	--	--	2.8	--	--	2.8
E	Exxon (Baton Rouge, LA)	--	--	10.0	--	--	10.0
E	Socal (Pascagoula, MS)	--	--	40.0	--	--	40.0
E	Tesoro (Carizzo Springs, TX)	--	--	12.0	--	--	12.0
E	South Hampton (Silsbee, TX)	--	--	12.5	--	--	12.5
R	Wickett Ref. (Wickett, TX)	--	--	8.0	--	--	8.0
N	Louisiana Land & Exploration (Mobile, AL)	--	--	30.0	--	--	30.0
E	Hunt (Tuscalossa, AL)	--	--	12.0	--	--	12.0
E	Thriftway (Graham, TX)	--	--	0.3	--	--	0.3

Source: Trends in Refining Capacity and Utilizations, FEA

TABLE A-4

NEW REFINERIES, EXPANSIONS AND REACTIVATIONS SCHEDULED IN THE UNITED STATES BY PAD DISTRICTS (Cont.)

(BPCD x 10^3 of Crude Distillation)

Type Code*	Company/Location	PAD Districts					Total
		I	II	III	IV	V	
E	Conoco (Commerce City, CO)	--	--	--	2.0	--	2.0
N	V-1 Oil Co. (Glenrock, WY)	--	--	--	1.0	--	1.0
E	Conoco (Paramount, CA)	--	--	--	--	11.0	11.0
E	U.S. Oil & Refining (Tacoma, WA)	--	--	--	--	3.0	3.0
N	United Ind. (Tacoma, WA)	--	--	--	--	1.0	1.0
E	HIRI (Oahu, HI)	--	--	--	--	15.0	15.0
Total (1975)		84.0	105.2	163.2	3.0	30.0	385.4
	1976						
E	Exxon (Bayway, NJ)	20.0	--	--	--	--	20.0
E	Midland Coop. (Cushing, OK)	--	16.0	--	--	--	16.0
E	American Petrofina (Port Arthur, TX)	--	--	34.0	--	--	34.0
E	Arco (Houston, TX)	--	--	92.0	--	--	92.0
N	Ecol (Garyville, LA)	--	--	200.0	--	--	200.0
N	Gulf Energy Refining (Luling, TX)	--	--	30.0	--	--	30.0
E	Champlin (Corpus Christi, TX)	--	--	60.0	--	--	60.0
E	Exxon (Baton Rouge, LA)	--	--	5.0	--	--	5.0
E	J & W Refining (Tucker, TX)	--	--	5.1	--	--	5.1
E	Thagard Oil (SouthGate, CA)	--	--	--	--	3.3	3.3
N	California Oil Pur. (Ventura County, CA)	--	--	--	--	15.0	15.0
E	Socal (Richmond, CA)	--	--	--	--	175.0	175.0
E	Socal (El Segunda, CA)	--	--	--	--	175.0	175.0
N	Pima (Phoenix, AZ)	--	--	--	--	3.3	3.3
	Misc. Net Additions	22.0	54.0	85.0	7.0	32.0	200.0
Total (1976)		42.0	70.0	511.1	7.0	403.6	1,033.7

TABLE A-4

NEW REFINERIES, EXPANSIONS AND REACTIVATIONS SCHEDULED IN THE UNITED STATES BY PAD DISTRICTS (Cont.)
(BPCD x 10^3 of Crude Distillation)

Type Code*	Company/Location	PAD Districts					Total
		I	II	III	IV	V	
	1977						
E	Exxon (Baytown, TX)	--	--	250.0	--	--	250.0
N	Energy Co. of Alaska (Fairbanks, AK)	--	--	--	--	15.0	15.0
	Misc. Net Additions	22.0	54.0	85.0	7.0	32.0	200.0
Total (1977)		22.0	54.0	335.0	7.0	47.0	465.0
	1978						
E	Texaco (Convent, LA)	--	--	200.0	--	--	200.0
N	Dow (Freeport, TX)	--	--	200.0	--	--	200.0
	Misc. Net Additions	22.0	54.0	85.0	7.0	32.0	200.0
Total (1978)		22.0	54.0	485.0	7.0	32.0	600.0
	1979						
	Pittston Co. (Eastport, ME)	250.0	--	--	--	--	250.0
	Misc. Net Additions	22.0	54.0	85.0	7.0	32.0	200.0
Total (1979)		272.0	54.0	85.0	7.0	32.0	450.0

*Type Code:

E--Expansion
N--New
R--Rehabilitation

TABLE A-5

REFINERIES PLANNED BUT NOT CONSTRUCTED DUE TO OPPOSITION ON ENVIRONMENTAL GROUNDS

Company	Location	Size-BPD	Final Action Blocking Project
Shell Oil Co.	Delaware Bay, DE	150,000	State reacted by legislature passing bill forbidding refineries in Coastal Area.
Fuels Desulfurization[1]	Riverhead, L.I.	200,000	City Council opposed project and would not change zoning.
Maine Clean Fuels[1]	South Portland, ME	200,000	City Council rejected proposal.
Maine Clean Fuels[1]	Searsport, ME	200,000	Maine Environmental Protection Board rejected proposal.
Georgia Refining Co.[1]	Brunswick, GA	200,000	Blocked through actions of Office of State Environmental Director.
Northeast Petroleum	Tiverton, RI	65,000	City Council rejected proposal.
Supermarine, Inc.	Hoboken, NJ	100,000	Hoboken Project withdrawn under pressure from environmental groups.
Commerce Oil	Jamestown Island, RI	50,000	Opposed by local organizations and contested in court.
Steuart Petroleum	Piney Point, MD	100,000	Rejected by St. Mary's County voters by referendum on July 23, 1974.
Olympic Oil Refineries, Inc.[2]	Durham, NH	400,000	Withdrawn after rejection by local referendum.
C. H. Sprague & Son	Newington, NH	50,000	Voted down in community vote on June 28, 1974.
Belcher Oil Co.	Manatee County, FL	200,000	Voted against in referendum September 10, 1974.

---

[1] Maine Clean Fuels and Georgia Refining Company are subsidiaries of Fuels Desulfurization and the refinery in question is the same in each case, so the capacity in barrels per day is not additive, but the incidents are independent and additive.

[2] Olympic is still considering other nearby sites.

company interviewed commented in detail on the scarcity of capital being one of the oil companies' major problems and one of the major inhibiting factors in refinery growth.

Manufacturing facilities, in general, have a shorter technical life now than a few decades ago, but the life is dependent upon the process. There are several operating crude units which were first constructed before World War II; they have, of course, undergone several modifications. On the other hand, "high" technology processes, such as hydrocracking, tend to have a shorter technical life, before some alterations, revamp, etc., have to be made. This is particularly true of the petrochemical complexes contained within several refineries.

The key variable in projecting the number and type of refineries is supply/demand projections. As a component of a recently published EPA report, ADL has just completed a major study in this area. Also, as part of this current task, several companies were interviewed to obtain their growth projections in several critical areas. The companies interviewed agreed in general that:

a. The gasoline growth rate between now and 1985 will be much less than for the 1965-1973 period. Some companies, in fact, project no growth at all; others about 2% per year maximum. The rationale used by all companies is based on improved fuel economy for automotive engines, coupled with a fairly rapid trend on the part of the consumer to purchase smaller (lighter) cars.

b. The major growth areas will be petrochemicals and fuel oils.

c. Jet fuel will probably not grow nearly as rapidly as was the case for 1965-1973.

ADL's product demand forecast has been described in detail in a recent EPA report (EPA 68-02-1332, Task Order No. 7). The methodology and assumptions used are also described in this report. The results of this forecast are shown in Table A-6, which shows the projected total U.S. energy demand for 1980 and 1985, and Table A-7, which projects the U.S. hydrocarbon product demand for 1977, 1980 and 1985. The latter table also shows the (estimated) consumption for the year 1975. As these tables show, the ADL forecasts are in general agreement with the views expressed by the companies interviewed for this task.

TABLE A-6

PROJECTION OF U.S. PRIMARY ENERGY SUPPLIES WITH OIL AS THE BALANCING FUEL
(Btu x 10^{12})

	Base Year	1980		1985	
	1972	High	Low	High	Low
Coal	12.3	19.6	19.8	23.6	21.5
Natural gas	25.1	23.9	23.9	24.7	24.7
Hydro	2.8	3.2	3.2	3.3	3.2
Nuclear	0.6	5.8	5.8	17.5	14.7
Non-conventional	-	0.1	0.1	0.2	0.2
Oil	32.5	42.5	39.2	44.7	40.9
TOTAL	73.3	95.1	92.0	114.0	105.2

Source: U.S. Bureau of Mines and Arthur D. Little, Inc. estimates

TABLE A-7

FORECAST OF U.S. PRODUCT DEMAND
(BPD x 10^3)

		1977		1980		1985	
	1975	High	Low	High	Low	High	Low
LPG & refinery gas	1.49	1.68	1.62	1.84	1.75	2.03	1.93
Naphtha & other	0.49	0.56	0.53	0.60	0.58	0.81	0.76
Gasoline	6.56	7.37	7.19	8.16	7.52	9.01	7.52
Kerosene & jet fuel	1.17	1.32	1.30	1.47	1.39	1.79	1.60
Distillate fuel oil	2.70	3.42	3.04	3.87	3.11	4.27	3.40
Residual fuel oil	2.52	3.14	2.79	4.03	3.24	3.02	2.75
Lubes, waxes, & coke	0.41	0.45	0.42	0.49	0.47	0.54	0.52
Asphalt	0.47	0.53	0.52	0.59	0.55	0.66	0.61
Refinery fuel & losses	1.28	1.43	1.41	1.62	1.53	1.78	1.69
TOTAL	17.09	19.90	18.82	22.67	20.14	23.91	20.78

APPENDIX B

PRESENT TECHNOLOGY

1. FEEDSTOCKS

There are literally thousands of different crude oils produced in the world today, a large number of which are processed in U.S. refineries. The crudes can be described by such terms as light, heavy, sweet, sour, domestic, foreign, lube stock, waxy, paraffinic, and the like. Any one refinery, other than the smallest and most simple, will process perhaps one or two dozen different crudes, in blends of varying composition. In some cases, crudes will be segregated (blocked) so that they can be processed separately. Yet another possibility is sending a particular crude to a dedicated process unit - e.g., to produce lube oils, petrochemical precursors, and the like.

The crude slate and process configuration impose limitations on the product mix that a refinery can produce. In the extreme, a change in crude type can downgrade refinery capacity and hence product output. Similarly, a refinery designed to run on sweet (low sulfur) crude may be incapable of processing sour (high sulfur) crude. There are two major reasons for this:

a. The sour crude is more corrosive and the metallurgy of the crude unit may not be adequate; and

b. The higher sulfur crude produces more pollutants (H_2S, SO_2, etc.) which the refinery may be incapable of removing without major alterations.

Another important factor is that each crude has different distillation properties, different yields of products, and products with different blending properties. To further make description of crude slates more difficult, two other factors must be considered.

a. Crude production and properties for a given oil field vary with time and location in the field. For example, not only has California crude production declined over the past few years, but the gravity of the crude oil has also become heavier. This tends to make the crude less valuable to a refinery because processing is more difficult and the yield of light products per barrel will go down.

b. Forecasting of crude production is extremely hard; yet to design new refineries, and make intelligent alterations to existing refineries, a major variable is the knowledge of the crude slate.

2. DESCRIPTION OF MAJOR REFINING PROCESSES

A modern, large, integrated refinery will use most of the following processes. (There are several types listed for some categories, such as catalytic cracking).

- Crude distillation;
- Vacuum distillation:
- Thermal processes, including gas oil cracking, vis-breaking, fluid coking, and delayed coking;
- Catalytic cracking, including fluid, thermoform, and houdriflow;
- Reforming, including catalytic (semiregenerative, conventional or bimetallic catalyst) and cyclic (both catalyst types);
- Hydrocracking, including distillate, residual, lubes, and other;
- Hydrorefining, including residual, heavy gas oil, residual, visbreaking, catalytic feed, and distillate;
- Hydrotreating, including reformer feed, naphtha, olefin/aromatics saturation, straight run distillate, lubes, and other distillates;
- Alkylation;
- Aromatic/Isomerization (BTX);
- Lube oil production; and
- Asphalt production.

In addition to these processes, there are units, such as desalters, sour water strippers, Claus plants, API separators, etc., within a refinery which do not process crude directly, but are vital components for pollution control or feed preparation. Also, a refinery will contain a large cooling tower with attendant blowdown water and washer treating facilities, plus a boiler for steam generation and perhaps power generation.

For process description of all of these refinery technologies, the reader is referred to the Refining Handbook Issue of Hydrocarbon Processing (1974).

3. COST FACTORS

In developing the cluster models referred to in Appendices A and C, a complete characterization of the economics for the processes listed above was made. A detailed description of these cost factors is contained in Appendix H (pp. H-37, 42) of our report to EPA on impact of lead additive regulation. The comparative economics in the present study were developed on an incremental basis and did not require the evaluation of economics for every process in the refinery.

APPENDIX C

BASE LINE ENERGY USE

1. REFINERY SIMULATION

The U.S. refining industry is composed of nearly 300 individual refineries scattered throughout the country, each characterized by a unique capacity, processing configuration, and product distribution. There are, however, logical regional groupings of major refineries with similar crude supply patterns, processing configurations, and product outputs. Therefore, the cluster model approach was used in developing base lines for this study. The existing U.S. refinery industry was simulated by the average operation of three similar refineries located in each of three selected regions, namely, East Coast (PAD I), Gulf Coast (PAD III), and West Coast (PAD V).

The selection of three representative refineries within each of the selected regions was accomplished with the assistance of the API/NPRA Task Force cooperating in a previous ADL study for EPA. (The Impact of Lead Additive Regulations on the Petroleum Refining Industry Contract No. 68-02-1332, T.O. No. 7). The most important criteria guiding the selection of these cluster models were: (1) each cluster model was to represent, as closely as possible, a realistic mode of operation, in that processing units were to be of normal commercial size, and that plants would be allowed normal flexibility in regard to raw material selection and product mix; (2) the cluster model crude slate, processing configurations, and product outputs were to bracket, as best as possible, those variations peculiar to each geographic region.

The final selection of refineries to be represented by the cluster models is shown in Table C-1.

PAD District I was simulated by three refineries in the Philadelphia-New Jersey area with capacities ranging from 160,000 to 255,000 BPD. In PAD District III, which represents about 40% of the U.S. refining capacity, differing types of refinery configurations can be identified. The Texas Gulf cluster is typified by a crude capacity exceeding 300,000 BPD and heavy involement in petrochemicals, lubes, and other specialty products. The Louisiana Gulf Coast cluster represents refineries between 174,000 and 240,000 BPD and which process a single source of sweet crude. Since the olefin industry was one of the 13 industry segments being studied, we chose the Louisiana cluster to represent the Gulf Coast region to avoid redundancy. PAD District V was simulated by a West Coast cluster model and was represented by refineries in the Southern California area.

TABLE C-1

REFINERIES SIMULATED BY CLUSTER MODELS

PAD District	Cluster Identification	Refineries Simulated	1973 Crude Capacity, 10^3 BPCD
I	East Coast	Arco - Philadelphia, Pa.	160.0
		Sun Oil - Marcus Hook, Pa.	163.0
		Exxon - Linden, New Jersey	255.0
III	Texas Gulf	Exxon - Baytown, Texas	350.0
		Gulf Oil - Port Arthur, Texas	312.1
		Mobil - Beaumont, Texas	335.0
III	Louisiana Gulf	Gulf Oil - Alliance, La.	174.0
		Shell Oil - Norco, La.	240.0
		Cities Service - Lake Charles, La.	240.0
V	West Coast	Mobil - Torrance, California	123.5
		Arco - Carson City, California	165.0
		Socal - El Segundo, California	220.0

After completion of the cluster refinery modeling, an extensive calibration/validation effort was undertaken by ADL with the assistance of the Bureau of Mines, Environmental Protection Agency and the API/NPRA Task Force. A complete discussion of the calibration/validation effort is contained in the referenced report. Only the highlights of this effort will be summarized here.

The annual refining surveys published in The Oil and Gas Journal were used as the basic reference source for determining the cluster model processing configurations, allowing simulation of those refineries listed in Table C-1. This source also provided the processing unit capacity available in these cluster refineries, used to limit the available capacity in the cluster models.

The input and output data required for each of these cluster models were obtained from the Bureau of Mines as 1973 annual data, for the aggregate of the three specific refineries comprising each individual cluster model (Table C-1). These data included the crude oil and other raw materials fed to the refineries, broken down by individual state or origin for domestic crudes and by country of origin for foreign sources. These data also included statistics on fuel consumed for all purposes in the refineries for the year 1973. Also included were all petroleum products from the cluster refineries for this base year.

The EPA and the API/NPRA Task Force obtained from each individual oil company the average gasoline grade distribution for the calendar year 1973 and the associated octane levels and lead levels for each grade. The EPA averaged their data to obtain information representing the cluster models, and supplied these data to ADL. Also obtained were total gasoline volumes and average sulfur contents as supplied by the individual companies and compiled by the EPA. Other key product specifications were obtained in a similar fashion and used in the calibration runs.

Four main areas were considered to compare the degree of calibration to the cluster models. These were: (1) overall refinery material balance (i.e., volume of the crude intake required to balance specified product demands and internal fuel requirements), (2) refinery energy consumption, (3) processing configuration, throughputs and operating severities, and (4) key product properties (e.g., gasoline clear pool octanes, lead levels, etc.). The refinery configurations and capacities are shown in Figures C-1, C-2, and C-3 for the East Coast, Gulf Coast, and West Coast calibration cases.

The refinery characteristics associated with the calibration cases are presented in Tables C-2 through C-7. Crude intakes and product slates for the three representative refineries are summarized in Tables C-2, C-3, and C-4. The refinery fuel balances are also shown in these tables. Although the basic refinery input/output data are based upon 1973 Bureau of Mines statistics, the information generated is also representative of operations until mid-1974, as little change occurred in the industry over that period.

A petroleum refinery is comprised of many integrated processing steps. The energy consumed by each processing unit within the refinery is summarized in Tables C-5, C-6, and C-7. The capacity of each unit process is also indicated. Purchased energy in the form of fuel, steam, and electricity is tabulated at the bottom of each table.

The major advantage of the calibrated models is that they can be used to determine refinery volume, energy and sulfur balances in future time, based on projected changes in crude intakes, product demands, and process configuration.

2. STUDY BASE LINE

There are certain established energy-related influences that will cause fundamental changes in petroleum refining operations, and thereby significantly alter refinery energy-consumption profiles within the next decade. Two important influences will be the phase-out of natural gas as refinery fuel and the phase-out of tetraethyl (TEL) lead in the gasoline pool. Our forecasts indicate that by 1985 both of these trends will have been accomplished. The implications of these trends should be quantified and taken into account in establishing a base line for comparing future process changes within the refining industry. Therefore, a 1985 refinery configuration and energy consumption profile was selected as the appropriate base line for comparing process changes within this industry. This approach reduces the aberrations associated with declining natural gas and TEL use and provides a more representative base line against which future process changes can be compared.

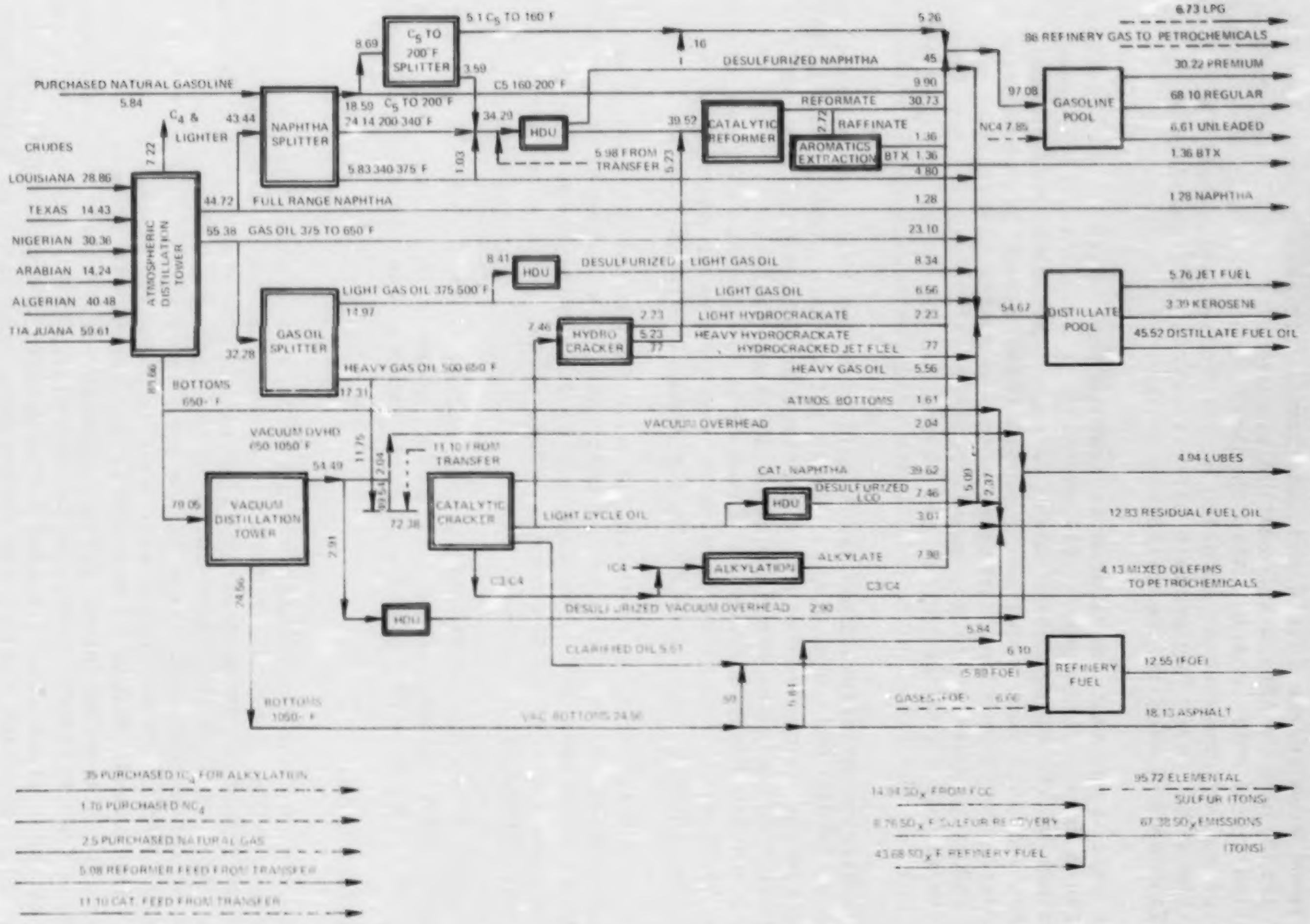

CRUDES
LOUISIANA 28.86
TEXAS 14.43
NIGERIAN 30.36
ARABIAN 14.24
ALGERIAN 40.48
TIA JUANA 59.61
PURCHASED NATURAL GASOLINE
5.84
ATMOSPHERIC DISTILLATION TOWER
7.22
C_4 & LIGHTER
43.44
NAPHTHA SPLITTER
8.69
C_5 TO 200°F SPLITTER
5.1 C_5 TO 160°F
3.59
C5 160-200°F
18.59 C_5 TO 200°F
24.14 200-340°F
5.83 340-375°F
1.03
34.29
HDU
5.98 FROM TRANSFER
5.23
.16
DESULFURIZED NAPHTHA
39.52
CATALYTIC REFORMER
REFORMATE
2.72
RAFFINATE
AROMATICS EXTRACTION
BTX
5.26
45
9.90
30.73
1.36
1.36
4.80
44.72
FULL RANGE NAPHTHA
1.28
55.38
GAS OIL 375 TO 650°F
23.10
8.41
HDU
DESULFURIZED LIGHT GAS OIL
8.34
LIGHT GAS OIL 375-500°F
14.97
LIGHT GAS OIL
6.56
32.28
GAS OIL SPLITTER
7.46
HYDRO CRACKER
2.23
5.23
.77
LIGHT HYDROCRACKATE
2.23
HEAVY HYDROCRACKATE
HYDROCRACKED JET FUEL
.77
HEAVY GAS OIL 500-650°F
17.31
HEAVY GAS OIL
5.56
85.66
BOTTOMS 650+°F
ATMOS. BOTTOMS
1.61
VACUUM OVERHEAD
2.04
VACUUM OVHD 650-1050°F
11.75
2.04
54.49
79.05
VACUUM DISTILLATION TOWER
99.54
72.38
11.10 FROM TRANSFER
CATALYTIC CRACKER
CAT. NAPHTHA
39.62
HDU
DESULFURIZED LCO
7.46
LIGHT CYCLE OIL
3.61
IC4
ALKYLATION
ALKYLATE
7.98
C3/C4
C3/C4
2.91
HDU
DESULFURIZED VACUUM OVERHEAD
2.90
5.00
2.37
5.84
CLARIFIED OIL 5.61
24.56
BOTTOMS 1050+°F
VAC BOTTOMS 24.56
.59
5.81
6.10
(5.89 FOE)
GASES (FOE)
6.66
REFINERY FUEL
97.08
NC4 7.85
GASOLINE POOL
6.73 LPG
.86 REFINERY GAS TO PETROCHEMICALS
30.22 PREMIUM
68.10 REGULAR
6.61 UNLEADED
1.36 BTX
1.28 NAPHTHA
54.67
DISTILLATE POOL
5.76 JET FUEL
3.39 KEROSENE
45.52 DISTILLATE FUEL OIL
4.94 LUBES
12.83 RESIDUAL FUEL OIL
4.13 MIXED OLEFINS TO PETROCHEMICALS
12.55 (FOE)
18.13 ASPHALT
.35 PURCHASED IC_4 FOR ALKYLATION
1.76 PURCHASED NC_4
2.5 PURCHASED NATURAL GAS
5.08 REFORMER FEED FROM TRANSFER
11.10 CAT. FEED FROM TRANSFER
14.94 SO_x FROM FCC
8.76 SO_x F. SULFUR RECOVERY
43.68 SO_x F. REFINERY FUEL
95.72 ELEMENTAL SULFUR (TONS)
67.38 SO_x EMISSIONS (TONS)

Figure C-1. East Coast Cluster Calibration

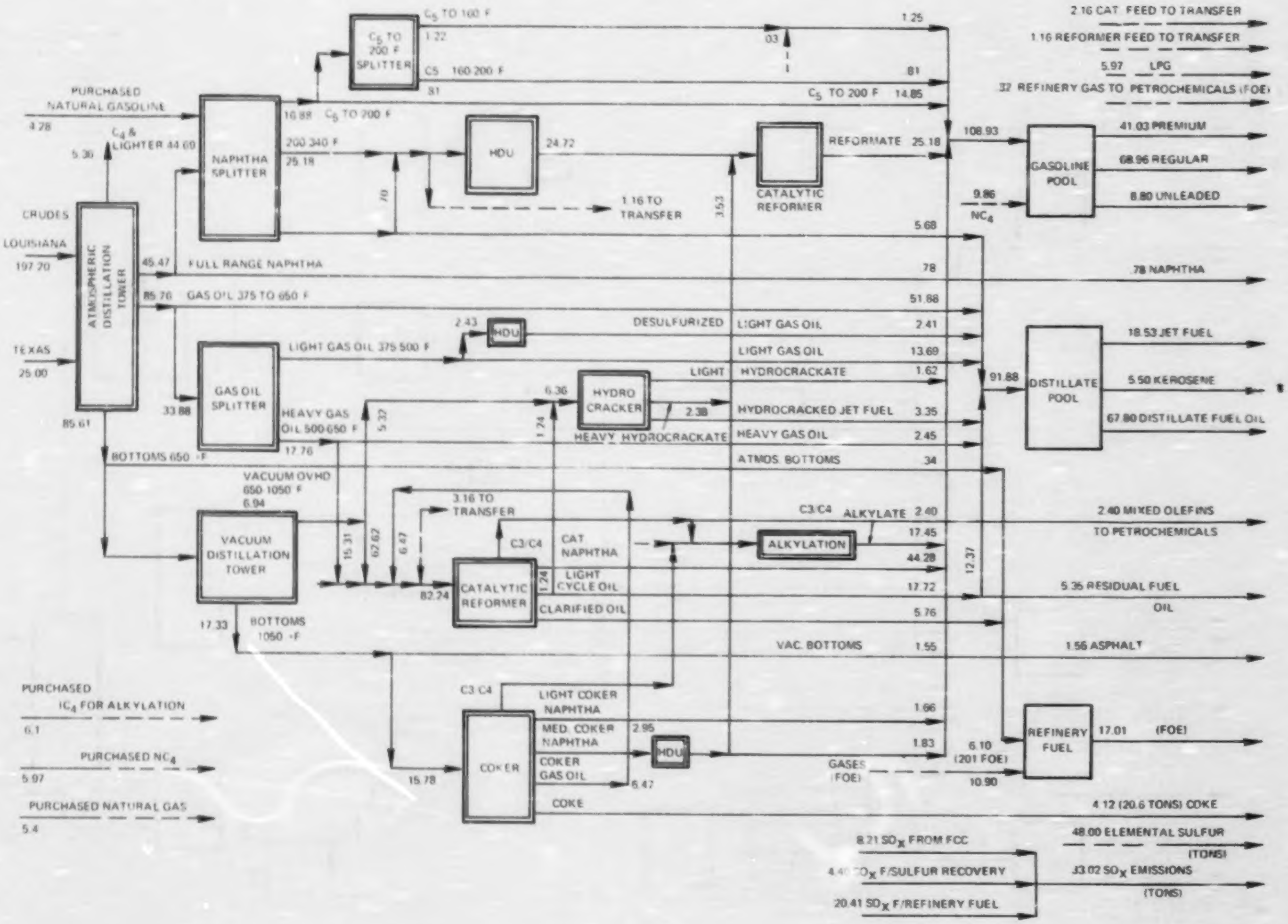

PURCHASED NATURAL GASOLINE
4.28
CRUDES
LOUISIANA
197.20
TEXAS
25.00
ATMOSPHERIC DISTILLATION TOWER
C_4 & LIGHTER 44.69
5.36
NAPHTHA SPLITTER
16.88
C_6 TO 200 F
C_5 TO 200 F SPLITTER
C_5 TO 160 F
1.22
C5 160-200 F
.81
200-340 F
25.18
.70
HDU
24.72
1.16 TO TRANSFER
3.53
CATALYTIC REFORMER
REFORMATE 25.18
1.25
.03
.81
C_5 TO 200 F 14.85
108.93
9.86
NC_4
GASOLINE POOL
41.03 PREMIUM
68.96 REGULAR
8.80 UNLEADED
2.16 CAT FEED TO TRANSFER
1.16 REFORMER FEED TO TRANSFER
5.97 LPG
.32 REFINERY GAS TO PETROCHEMICALS (FOE)
45.47
FULL RANGE NAPHTHA
5.68
.78
.78 NAPHTHA
85.76
GAS OIL 375 TO 650 F
51.88
33.88
GAS OIL SPLITTER
LIGHT GAS OIL 375-500 F
2.43
HDU
DESULFURIZED LIGHT GAS OIL
2.41
LIGHT GAS OIL
13.69
HEAVY GAS OIL 500-650 F
17.76
5.37
6.36
1.24
HYDRO CRACKER
LIGHT HYDROCRACKATE
1.62
2.38
HYDROCRACKED JET FUEL
3.35
HEAVY HYDROCRACKATE
HEAVY GAS OIL
2.45
91.88
DISTILLATE POOL
18.53 JET FUEL
5.50 KEROSENE
67.80 DISTILLATE FUEL OIL
85.61
BOTTOMS 650 +F
ATMOS. BOTTOMS
.34
VACUUM OVHD 650-1050 F
6.94
VACUUM DISTILLATION TOWER
15.31
67.62
6.47
3.16 TO TRANSFER
82.24
CATALYTIC REFORMER
C3/C4
CAT NAPHTHA
1.24
LIGHT CYCLE OIL
CLARIFIED OIL
C3/C4
ALKYLATE
2.40
ALKYLATION
17.45
44.28
17.72
5.76
12.37
2.40 MIXED OLEFINS TO PETROCHEMICALS
5.36 RESIDUAL FUEL OIL
17.33
BOTTOMS 1050 +F
VAC. BOTTOMS
1.55
1.56 ASPHALT
15.78
COKER
C3 C4
LIGHT COKER NAPHTHA
MED. COKER NAPHTHA
2.95
HDU
COKER GAS OIL
6.47
COKE
1.66
1.83
GASES (FOE)
6.10
(201 FOE)
10.90
REFINERY FUEL
17.01 (FOE)
4.12 (20.6 TONS) COKE
PURCHASED IC_4 FOR ALKYLATION
6.1
PURCHASED NC_4
5.97
PURCHASED NATURAL GAS
5.4
8.21 SO_x FROM FCC
4.40 SO_x F/SULFUR RECOVERY
20.41 SO_x F/REFINERY FUEL
48.00 ELEMENTAL SULFUR (TONS)
33.02 SO_x EMISSIONS (TONS)

Figure C-2. Louisiana Gulf Cluster Model Calibration

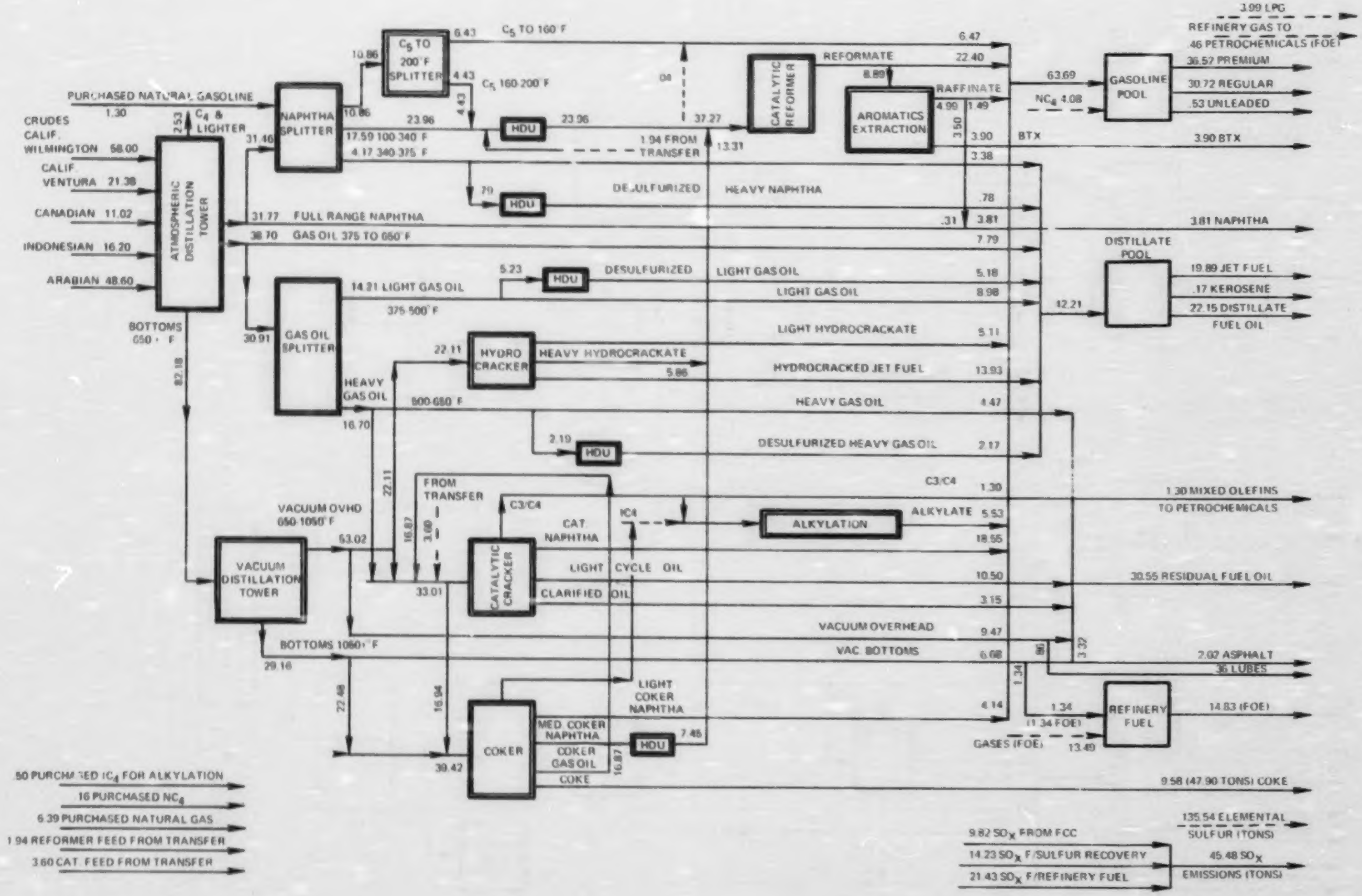

PURCHASED NATURAL GASOLINE
1.30
CRUDES
CALIF.
WILMINGTON 58.00
CALIF.
VENTURA 21.38
CANADIAN 11.02
INDONESIAN 16.20
ARABIAN 48.60
2.53
C_4 & LIGHTER
ATMOSPHERIC DISTILLATION TOWER
BOTTOMS 650 + F
82.18
31.46
NAPHTHA SPLITTER
10.86
10.86
C_5 TO 200°F SPLITTER
6.43
C_5 TO 160°F
4.43
C_5 160-200°F
4.43
23.96
17.59 100-340°F
4.17 340-375°F
HDU
23.96
.04
37.27
1.94 FROM TRANSFER
13.31
CATALYTIC REFORMER
.79
HDU
DESULFURIZED
HEAVY NAPHTHA
31.77 FULL RANGE NAPHTHA
38.70 GAS OIL 375 TO 650°F
30.91
GAS OIL SPLITTER
14.21 LIGHT GAS OIL
375-500°F
5.23
HDU
DESULFURIZED
LIGHT GAS OIL
LIGHT GAS OIL
22.11
HYDRO CRACKER
LIGHT HYDROCRACKATE
HEAVY HYDROCRACKATE
5.86
HYDROCRACKED JET FUEL
HEAVY GAS OIL
500-650°F
16.70
HEAVY GAS OIL
2.19
HDU
DESULFURIZED HEAVY GAS OIL
22.11
FROM TRANSFER
16.87
3.60
VACUUM OVHD 650-1050°F
53.02
VACUUM DISTILLATION TOWER
33.01
CATALYTIC CRACKER
C3/C4
CAT. NAPHTHA
LIGHT CYCLE OIL
CLARIFIED OIL
IC4
ALKYLATION
C3/C4
ALKYLATE
BOTTOMS 1050+°F
29.16
22.48
16.94
39.42
COKER
MED. COKER NAPHTHA
COKER GAS OIL
COKE
16.87
HDU
7.46
LIGHT COKER NAPHTHA
VACUUM OVERHEAD
VAC. BOTTOMS
REFORMATE
8.89
AROMATICS EXTRACTION
RAFFINATE
4.99
1.49
3.50
3.90
BTX
6.47
22.40
3.38
.78
.31
3.81
7.79
5.18
8.98
5.11
13.93
4.47
2.17
1.30
5.53
18.55
10.50
3.15
9.47
6.68
4.14
.96
3.32
1.34
63.69
NC_4 4.08
GASOLINE POOL
3.99 LPG
REFINERY GAS TO
.46 PETROCHEMICALS (FOE)
36.57 PREMIUM
30.72 REGULAR
.53 UNLEADED
3.90 BTX
3.81 NAPHTHA
DISTILLATE POOL
12.21
19.89 JET FUEL
.17 KEROSENE
22.15 DISTILLATE FUEL OIL
1.30 MIXED OLEFINS TO PETROCHEMICALS
30.55 RESIDUAL FUEL OIL
2.02 ASPHALT
.36 LUBES
1.34
(1.34 FOE)
GASES (FOE)
13.49
REFINERY FUEL
14.83 (FOE)
9.58 (47.90 TONS) COKE
.50 PURCHASED IC_4 FOR ALKYLATION
.16 PURCHASED NC_4
6.39 PURCHASED NATURAL GAS
1.94 REFORMER FEED FROM TRANSFER
3.60 CAT. FEED FROM TRANSFER
9.82 SO_x FROM FCC
14.23 SO_x F/SULFUR RECOVERY
21.43 SO_x F/REFINERY FUEL
135.54 ELEMENTAL SULFUR (TONS)
45.48 SO_x EMISSIONS (TONS)

Figure C-3. West Coast Cluster Model Calibration

TABLE C-2

SUMMARY OF EAST COAST REFINERY OPERATION (1974)

Refinery Location East Coast Year 1974

Nominal Capacity 200 10^3 BPD

Refinery Intakes	10^3 BPD
Crudes	
1. Louisiana	28.9
2. Nigerian	30.4
3. Arabian Lt.	14.2
4. W. Texas Sour	14.4
5. Algerian	40.5
6. Tiajuana	59.6
Purchased IC_4	0.4
Purchased NC_4	1.8
Purchased nat. gasoline	5.8
Purchased nat. gas	2.5
Cat. feed from storage	11.1
Reform. feed from storage	6.0
Total	215.6

Refinery Products	10^3 BPD
Regular gasoline	68.1
Premium gasoline	30.2
Unleaded gasoline	6.6
Naphtha	-
No. 2 distillate	45.5
Jet fuel	5.8
Kerosene	3.4
Low S F.O.	-
Residual F.O.	12.8
Asphalt	18.1
Lube stocks	4.9
Coke	-
BTX	1.4
Mixed olefins (FOE)	4.1
Total	200.9

Refinery Fuel Balance

Sources	FOE[1]/day
Isomerization	
Atm. distillation	10
Cat. reforming	2760
Cat. cracking	2010
Hydrocracking	-30
Coker gas	-
Imported natural gas	2500
LPG or fuel oil (blend)	7210
Total	14,520

Consumers	FOE/day
Hydrogen production	
feedstock	966
fuel	717
Fuel to petrochemicals	860
Steam generation	1793
Process heater	10,184
Total	14,520

---

[1] FOE = 6.3×10^6 Btu (HHV)

TABLE C-3

SUMMARY OF GULF COAST REFINERY OPERATION

Refinery Location Gulf Coast Year 1974
Nominal Capacity 200 10^3 BPD

Refinery Intakes	10^3 BPD
Crudes	
1. Louisiana Sweet	197
2. W. Texas Sour	25
Purchased IC_4	6
Purchased NC_4	6
Purchased nat. gasoline	4.3
Purchased nat. gas	5.4
Cat. feed from storage	--
Reform. feed from storage	--
Total	243.7

Refinery Products	10^3 BPD
Regular gasoline	69
Premium gasoline	41
Unleaded gasoline	8.8
Naphtha	67.8
No. 2 distillate	67.8
Jet fuel	18.5
Kerosene	5.5
Low S F.O.	5.4
Residual F.O.	--
Asphalt	1.6
Lube stocks	--
Coke	4.1
BTX	--
Mixed olefins (FOE)	2.4
Total	224.1

Refinery Fuel Balance

Sources	FOE[1]/day
Isomerization	
Atm. distillation	10
Cat. reforming	1,420
Cat. cracking	2,490
Hydrocracking	30
Coker gas	1,500
Imported natural gas	5,400
LPG or fuel oil (blend)	5,350
Total	16,200

Consumers	FOE/day
Hydrogen production	
feedstock	568
fuel	422
Fuel to petrochemicals	320
Steam generation	3,200
Process heaters	11,690
Total	16,200

---

[1] FOE = 6.3×10^6 Btu (HHV)

TABLE C-4

SUMMARY OF WEST COAST REFINERY OPERATION

Refinery Location West Coast Year 1974

Nominal Capacity 175 10^3 BPD

Refinery Intakes	10^3 BPD
Crudes	
1. Wilmington	58.0
2. Ventura	21.4
3. Canadian	11.0
4. Imported sour	48.6
5. Indonesia-Mines	16.2
Purchased IC_4	0.5
Purchased NC_4	0.2
Purchased nat. gasoline	1.3
Purchased nat. gas	6.4
Cat. feed from storage	3.6
Reform. feed from storage	1.9
Total	169.1

Refinery Products	10^3 BPD
Regular gasoline	30.7
Premium gasoline	36.5
Unleaded gasoline	0.5
Naphtha	
No. 2 distillate	22.2
Jet fuel	19.9
Kerosene	0.2
Low S F.O.	30.6
Residual F.O.	
Asphalt	2.0
Lube stocks	0.4
Coke	9.6
BTX	3.9
Mixed olefins (FOE)	1.3
Total	157.8

Refinery Fuel Balance

Sources	FOE[1]/day
Isomerization	
Atm. distillation	10
Cat. reforming	2,460
Cat. cracking	880
Hydrocracking	90
Coker gas	3,750
Imported natural gas	6,390
LPG or fuel oil (blend)	2,140
Total	15,720

Consumers	FOE/day
Hydrogen production	
feedstock	1,739
fuel	1,289
Fuel to petrochemicals	130
Steam generation	3,373
Process heaters	9,189
Total	15,720

---

[1] FOE = 6.3×10^6 Btu (HHV)

TABLE C-5

SUMMARY OF EAST COAST REFINERY ENERGY CONSUMPTION (1974)

Process Unit	Unit Capacity BPD x 10^3	Refinery Fuel 10^6 Btu/day	Steam 10^3 lb/day	Electricity 10^3 kWh/day
Atmospheric distillation	188	23,732	8,098	235.4
Vacuum distillation	79.1	5,097	2,425	16.2
Cat. cracking	72.4	8,694	68	138.0
Alkylation (product basis)	8.0	4,624	220	27.2
Isomerization	-	-	-	-
Cat. reforming	39.5	14,421	46	150.3
Hydrocracking	7.5	1,399	111	74.0
Coking	-	-	-	-
H_2 production (feed & fuel)	23 10^6 scfpd	10,603	23	4.0
Sulfur recovery[1]	93.8 tpd	1,688	447	8.9
Naphtha splitter	43.4	384	-	-
Atm. G.O. splitter	32.3	410	-	-
Aromatics extraction	2.7	36	1,088	5.4
Naphtha desulfurization[2]	34.3	2,363	375	39.0
Gas oil desulfurization[3]	8.4	712	113	14.4
Cycle oil desulfurization	7.6	560	89	11.6
Coker naphtha desulfurization	-	-	-	-
Steam boilers	-	11,296	(7,472)	3.7
Total Consumed		86,069	5,631	713.7
Purchased Energy		15,750	5,631	713.7

---

[1] Sulfur plant steam used to boil up amine stripper.
[2] Includes light and heavy naphtha.
[3] Includes light and heavy gas oil.

TABLE C-6

SUMMARY OF GULF COAST REFINERY ENERGY CONSUMPTION (1974)

Process Unit	Unit Capacity BPD x 10^3	Refinery Fuel 10^6 Btu/day	Steam 10^3 lb/day	Electricity 10^3 kWh/day
Atmospheric distillation	222.2	27,997	9,565	277.8
Vacuum distillation	85.3	5,368	2,558	17.1
Cat. cracking	82.2	12,764	82	172.1
Alkylation (product basis)	17.5	10,994	524	64.6
Isomerization	--	--	--	--
Cat. reforming	28.3	8,915	28	84.8
Hydrocracking	6.6	1,172	96	60.3
Coking	15.8	2,986	474	23.7
H_2 production (feed & fuel)	13.6 10^6 scfpd	6,243	14	2.3
Sulfur recovery[1]	61.7 tpd	806	213	4.3
Naphtha splitter	44.7	284	--	--
Atm. G.O. splitter	33.9	428	--	--
Aromatics extraction	--	--	--	--
Naphtha desulfurization[2]	24.7	1,556	247	24.7
Gas oil desulfurization[3]	2.4	151	24	2.9
Cycle oil desulfurization	--	--	--	--
Coker naphtha desulfurization	3.0	189	30	2.9
Steam boilers	--	20,172	(13,340)	6.7
Total Consumed		100,025	505	744.2
Purchased Energy		34,020	505	744.2

---

[1] Sulfur plant steam used to boil up amine stripper.
[2] Includes light and heavy naphtha.
[3] Includes light and heavy gas oil.

TABLE C-7

SUMMARY OF WEST COAST REFINERY ENERGY CONSUMPTION (1974)

Process Unit	Unit Capacity BPD x 10^3	Refinery Fuel 10^6 Btu/day	Steam 10^3 lb/day	Electricity 10^3 kWh/day
Atmospheric distillation	155.2	19,555	6,674	194.0
Vacuum distillation	82.2	5,179	2,464	16.4
Cat. cracking	35.0	4,410	35	70.0
Alkylation (product basis)	5.5	3,484	166	20.5
Isomerization				
Cat. reforming	37.3	12,153	37	120.2
Hydrocracking	22.1	3,900	322	199.0
Coking	39.4	3,200	508	59.1
H_2 production (feed & fuel)	41.6 10^6 scfpd	19,076	42	7.1
Sulfur recovery[1]	139.5 tpd	2,615	691	13.8
Naphtha splitter	31.5	258	-	-
Atm. G.O. splitter	30.9	391	-	-
Aromatics extraction	-	284	3,560	17.8
Naphtha desulfurization[2]	24.7	1,525	242	24.8
Gas oil desulfurization[3]	7.4	466	74	8.9
Cycle oil desulfurization	-	-	-	-
Coker naphtha desulfurization	7.4	466	74	7.4
Steam boilers		21,250	(15,570)	7.8
Total Consumed		98,212	(682)	766.8
Purchased Energy		40,257	682	766.8

---

[1] Sulfur plant steam used to boil up amine stripper.

[2] Includes light and heavy naphtha.

[3] Includes light and heavy gas oil.

The calibrated cluster models were utilized to characterize the 1985 operations with the above constraints in effect. Information on refinery intake and outputs, fuel balances, and energy consumption by unit process are summarized in Tables C-8 to C-10 and C-11 to C-13 for the 1985 time-frame. These base line scenarios assume no natural gas component in the refinery fuel and the production of only lead-free gasoline. The important shifts in refinery energy usage occurring over this period (1974-1985) are presented in Table C-14. In 1985, the total refinery fuel requirement must be furnished from internal energy sources. This is accomplished through increased consumption of refinery gas (methane/ethane) and blended fuel oil fractions. Restrictions were imposed on the refinery fuel oil sulfur level in each region. These restrictions comply with the existing legislated limits as outlined below:

Region	Maximum Sulfur, wt.%
East Coast	0.6
Louisiana Gulf	0.9
West Coast	0.7

As indicated in Table C-14, fuel and electric power consumption will be about 18% higher in 1985. Refinery energy consumption will generally be higher due primarily to catalytic reformer severity upgrading to meet unleaded gasoline RON values (see discussion in Chapter IV-C).

3. ENERGY CONSIDERATIONS

Historically, the refinery fuel balance was made up of purchased natural gas, refinery fuel gas, and oil. In the Gulf Coast and West Coast regions, natural gas furnished from one-third to one-half of the refinery fuel balance. The fuel consumed by a refinery in the processing of crude to finished product is in the order of 8 to 11% of crude runs. Since the management of the refinery fuel system is an integral part of the refinery operation, the allocation of intermediate product streams can have an effect on the final product slates. For example, switching large process heaters and boilers from gas to liquid fuels is technically feasible and the displaced $C_{\overline{2}}$ gas could be exported for higher priority uses. Consequently, changes in the refinery fuel blend can result in energy conservation in terms of form value.

Changes in processing, particularly upgrading, also have important energy implications. While these processes in general may result in an increase in overall refinery energy consumption, they produce higher valued light products from petroleum residues and thereby result in a form value benefit.

In summary, the operation of a refinery is moderately energy-intensive and the potential form value conservation is relatively high due to the ability to modify the product slate through management of the fuel system and modification of the process sequences. Reductions in the 6-10% of crude consumed as fuel are also possible through better heat recovery and "energy" auditing; however, these methods are excluded from the scope of this study.

TABLE C-8

SUMMARY OF EAST COAST REFINERY OPERATION
(1985)

Refinery Location East Coast Year 1985

Nominal Capacity 200 10^3 BPD

Refinery Intakes	10^3 BPD
Crudes	
1. Tiajuana	32.7
2. Nigerian	38.0
3. Arabian Light	90.4
4. Algerian	36.8
Purchased IC_4	0.2
Purchased NC_4	1.2
Purchased nat. gasoline	5.8
Purchased nat. gas	-
Cat. feed from storage	11.1
Reform. feed from storage	6.0
Total	222.2

Refinery Products	10^3 BPD
Regular gasoline	-
Premium gasoline	-
Unleaded gasoline	106.9
Naphtha	1.3
No. 2 distillate	47.0
Jet fuel	5.9
Kerosene	3.5
Low S F.O.	13.2
Residual F.O.	-
Asphalt	18.7
Lube stocks	5.1
Coke	-
BTX	1.4
Mixed olefins (FOE)	4.3
Total	207.3

Refinery Fuel Balance

Sources	FOE[1]/day
Isomerization	270
Atm. distillation	60
Cat. reforming	3,740
Cat. cracking	2,880
Hydrocracking	40
Coker gas	-
Imported natural gas	-
LPG or fuel oil (blend)	9,640
Total	16,630

Consumers	FOE/day
Hydrogen production	1,580
feedstock	907
fuel	673
Fuel to petrochemicals	890
Steam generation	1,920
Process heaters	12,240
Total	16,630

---

[1] FOE = 6.3×10^6 Btu (HHV)

TABLE C-9

SUMMARY OF GULF COAST REFINERY OPERATION
(1985)

Refinery Location Gulf Coast Year 1985

Nominal Capacity 200 10^3 BPD

Refinery Intakes	10^3 BPD	Refinery Products	10^3 BPD
Crudes		Regular gasoline	-
1. Texas Sour	25.7	Premium gasoline	-
2. Louisiana Sweet	192.3	Unleaded gasoline	112.5
		Naphtha	0.7
		No. 2 distillate	63.8
		Jet fuel	17.4
		Kerosene	5.2
Purchased IC_4	4.7	Low S F.O.	5.0
Purchased NC_4	4.2	Residual F.O.	-
Purchased nat. gasoline	3.0	Asphalt	1.5
Purchased nat. gas	-	Lube stocks	-
Cat. feed from storage	-	Coke	3.9
Reform. feed from storage	-	BTX	-
Total	229.9	Mixed olefins (FOE)	2.3
		Total	212.3

Refinery Fuel Balance

Sources	FOE[1]/day	Consumers	FOE/day
Isomerization	230	Hydrogen production	1,270
		feedstock	729
		fuel	541
Atm. distillation	10	Fuel to petrochemicals	300
Cat. reforming	2,880	Steam generation	3,250
Cat. cracking	2,160	Process heaters	12,760
Hydrocracking	30	Total	17,580
Coker gas	1,570		
Imported natural gas	-		
LPG or fuel oil (blend)	10,700		
Total	17,580		

---

[1] FOE = 6.3×10^6 Btu (HHV)

TABLE C-10

SUMMARY OF WEST COAST REFINERY OPERATION (1985)

Refinery Location West Coast Year 1985

Nominal Capacity 175 10^3 BPD

Refinery Intakes	10^3 BPD
Crudes	
1. California Wilmington	65.7
2. California Ventura	21.7
3. Alaskan-North Slope	76.8
Purchased IC_4	0.4
Purchased NC_4	0.1
Purchased nat. gasoline	0.9
Purchased nat. gas	-
Cat. feed from storage	3.6
Reform. feed from storage	1.9
Total	171.1

Refinery Products	10^3 BPD
Regular gasoline	-
Premium gasoline	-
Unleaded gasoline	70.3
Naphtha	3.9
No. 2 distillate	22.7
Jet fuel	20.4
Kerosene	0.2
Low S F.O.	-
Residual F.O.	31.3
Asphalt	2.1
Lube stocks	0.4
Coke	7.8
BTX	4.0
Mixed olefins (FOE)	1.3
Total	164.4

Refinery Fuel Balance

Sources	FOE[1]/day
Isomerization	60
Atm. distillation	50
Cat. reforming	3,100
Cat. cracking	1,100
Hydrocracking	110
Coker gas	3,080
Imported natural gas	-
LPG or fuel oil (blend)	11,690
Total	19,190

Consumers	FOE/day
Hydrogen production	3,750
feedstock	2,153
fuel	1,597
Fuel to petrochemicals	470
Steam generation	4,220
Process heaters	10,750
Total	19,190

---

[1] FOE = 6.3 x 10^6 Btu (HHV)

TABLE C-11

SUMMARY OF EAST COAST REFINERY ENERGY CONSUMPTION (1985)

Process Unit	Unit Capacity BPD x 10^3	Refinery Fuel 10^6 Btu/day	Steam 10^3 lb/day	Electricity 10^3 kWh/day
Atmospheric distillation	197.9	25,439	8,510	247.4
Vacuum distillation	68.7	4,568	2,174	14.5
Cat. cracking	62.2	14,515	58	138.2
Alkylation (product basis)	12.9	7,629	363	44.8
Isomerization	7.6	1,581	84	35.4
Cat. reforming	46.5	15,145	47	168.3
Hydrocracking	9.4	1,688	136	87.7
Coking	--	--	--	--
H_2 production (feed & fuel)	20.7 10^6 scfpcd	9,941	23	3.9
Sulfur recovery[1]	56.8 stpcd	1,676	443	8.9
Naphtha splitter	75.0	473	--	--
Atm. G.O. splitter	33.2	416	--	--
Aromatics extraction	2.8	88	1,122	5.6
Naphtha desulfurization[2]	49.6	3,219	511	51.1
Gas oil desulfurization[3]	15.9	677	197	23.9
Cycle oil desulfurization	4.1	--	--	--
Coker naphtha desulfurization	--	--	--	--
Steam boilers	--	12,096	(8,001)	4.0
Total Consumed		99,151	5,669	833.7
Purchased Energy		--	5,669	833.7

---

[1] Sulfur plant steam used to boil up amine stripper.

[2] Includes light and heavy naphtha.

[3] Includes light and heavy gas oil.

TABLE C-12

SUMMARY OF GULF COAST REFINERY ENERGY CONSUMPTION (1985)

Process Unit	Unit Capacity BPD x 10^3	Refinery Fuel 10^6 Btu/day	Steam 10^3 lb/day	Electricity 10^3 kWh/day
Atmospheric distillation	218.0	27,468	9,374	272.5
Vacuum distillation	75.8	4,775	2,273	15.2
Cat. cracking	80.2	12,795	80	166.6
Alkylation (product basis)	18.1	11,422	544	67.1
Isomerization	8.1	1.537	81	27.9
Cat. reforming	35.5	11,164	35	124.1
Hydrocracking	8.8	1,562	128	79.6
Coking	17.6	3,333	529	26.4
H_2 production (feed & fuel)	17.5 10^6 scfpcd	8,026	18	3.0
Sulfur recovery[1]	30.2 stpd	914	242	4.8
Naphtha splitter	60.4	359	--	--
Atm. G.O. splitter	36.8	466	--	--
Aromatics extraction	--	--	--	--
Naphtha desulfurization[2]	38.1	2,400	381	31.4
Gas oil desulfurization[3]	31.7	1,997	317	46.7
Cycle oil desulfurization	--	--	--	--
Coker naphtha desulfurization	3.1	189	30	3.0
Steam boilers	--	20,462	(13,532)	6.8
Total Consumed		108,869	500	875.1
Purchased Energy		--	500	875.1

---

[1] Sulfur plant steam used to boil up amine stripper.

[2] Includes light and heavy naphtha.

[3] Includes light and heavy gas oil.

TABLE C-13

SUMMARY OF WEST COAST REFINERY ENERGY CONSUMPTION (1985)

Process Unit	Unit Capacity BPD x 10^3	Refinery Fuel 10^6 Btu/day	Steam 10^3 lb/day	Electricity 10^3 kWh/day
Atmospheric distillation	164.2	20,689	7,060	205.2
Vacuum distillation	79.3	4,996	2,378	15.9
Cat. cracking	38.2	5,676	38	79.1
Alkylation (product basis)	7.9	5,002	238	29.4
Isomerization	5.1	964	51	18.4
Cat. reforming	38.0	11,939	38	133.1
Hydrocracking	27.4	4,830	397	246.4
Coking	33.0	6,237	990	49.5
H_2 production (feed & fuel)	51.5 10^6 scfpcd	23,600	52	8.8
Sulfur recovery[1]	85.4	2,583	684	13.7
Naphtha splitter	38.5	239	--	--
Atm. G.O. splitter	40.6	510	--	--
Aromatics extraction	12.7	397	5,068	25.3
Naphtha desulfurization[2]	29.8	1,877	298	29.8
Gas oil desulfurization[3]	14.9	980	156	18.7
Cycle oil desulfurization	7.8	445	71	9.2
Coker naphtha desulfurization	6.1	381	61	6.0
Steam boilers		26,580	(17,578)	8.8
Total Consumed		117,925	0	897.3
Purchased Energy		---	---	897.3

---

[1] Sulfur plant steam used to boil up amine stripper.
[2] Includes light and heavy naphtha.
[3] Includes light and heavy gas oil.

TABLE C-14

ANTICIPATED SHIFTS IN REFINERY ENERGY USAGE CAUSED BY NATURAL GAS AND TEL PHASE-OUT

	East Coast		Gulf Coast		West Coast	
	1974	1985	1974	1985	1974	1985
Crude Run, 10^3 BPD	188	198	222	218	155	164
Purchased Natural Gasoline, FOE/day	2,500	---	5,400	---	6,390	---
Refinery Fuel, FOE/day						
Gas	4,810	6,990	5,450	6,880	7,190	7,500
Oil (Blended)	7,210	9,640	5,350	10,700	2,140	11,690
TOTAL FUEL, FOE/day	14,520	16,630	16,200	17,580	15,720	19,190
Purchased Steam, 10^6 lb/day	5.6	5.7	0.5	0.5	0.7	0
Electricity, MWh/day	713.7	838.7	744.2	875.1	766.8	897.3

4. TYPICAL REFINERY OPERATING COST

Refinery operating costs can vary considerably depending upon the size, efficiency and age of a given facility. The simulation models applied in this study are more representative of large, modern refineries.

Typical operating costs for the 1985 base line refineries are summarized in Table C-15. The costs are presented in dollars per stream day based on constant 1975 dollars. Capital charges on existing investment were based on an estimated book balue of $1,000/bbl of daily capacity. Working capital was estimated on the basis of three months' supply of crude oil and products, with crude valued at $7.66/bbl and $12.50/bbl for domestic and imported, respectively.

TABLE C-15

CAPITAL INVESTMENTS AND OPERATING COSTS
(Constant 1975 dollars)

	East Coast	Gulf Coast	West Coast
Capital Investments ($\$10^9$) (1st Q 1975 Basis)			
Reforming: existing capacity	---	1.8	5.2
severity upgrade	6.0	1.8	0.1
new capacity	---	0.8	---
Hydrocracking: existing capacity	---	---	---
new capacity	0.1	---	---
Isomerization: once through upgrading	1.9	7.3	4.8
new capacity	5.3	---	---
Alkylation: new capacity	0.4	0.7	0.4
Light naphtha desulfurization: new capacity	1.4	2.6	1.6
Subtotals	15.1	15.0	12.1
Offsites and working capital at 40%	6.0	6.0	4.8
Subtotals	21.1	21.0	16.9
Approximate book value	200.0	200.0	170.0
Operating Costs ($\$10^3$/day)			
Purchased steam	21.9	1.6	---
Electricity	18.0	12.3	16.2
Cooling water	16.6	17.3	12.5
Maintenance	30.8	31.5	29.6
Manpower	53.8	44.5	71.0
Catalysts and chemicals	20.1	16.6	16.5
Fixed Costs			
Depreciation	40.3	40.3	34.3
Capital charges[1]	480.4	434.5	359.6
TOTAL	681.9	598.6	539.7

---

[1] On investment and working capital.

APPENDIX D

CURRENT POLLUTION PROBLEMS AND EFFECTIVENESS OF AVAILABLE POLLUTION CONTROL TECHNOLOGY
(Base Line Characterization)

1. WATER POLLUTION

To assess the water pollution implications of anticipated energy-saving process changes within the petroleum refining industry, it was deemed necessary to first delineate the base case petroleum refinery in terms of the sources and characteristics of wastewater and wastewater treatment technology and economics.

a. Sources of Wastewater

The major use of water in petroleum refining is for steam generation and heat transfer. The volume of water coming in direct contact with process streams is small when compared with water for indirect cooling and heat transfer. Nevertheless, almost every one of the major refining operations produces a wastewater stream containing various pollutants. For example, ballast water pumped out from incoming tankers is contaminated with previously stored petroleum products. Tank farms used for refinery product storage produce wastewater streams due to stormwater run-off contacting petroleum-contaminated exposed areas. Within the confines of the refinery itself, there are numerous processing steps in which steam, condensate, or cooling water comes in contact with petroleum and/or petroleum products. Superimposed over these major, rather continuous, wastewater generation points are countless leaks and spills which eventually drain into the central refinery sewer system. Stormwater run-off from process areas is another significant source of wastewater.

All petroleum refineries use large amounts of non-contact cooling water. Due to the large volumes required, it is an almost universal practice to install a cooling tower circuit and reuse most of the water. To prevent the buildup of naturally occurring salts in the cooling water circuit, it is necessary to purge a portion of the total cooling water flow. Since it is a common practice to use corrosion inhibitors in the cooling circuit, cooling tower blowdown will contain such substances.

A number of attempts have been made to quantify both the flow rates and characteristics of wastewater associated with the various refining operations. ("The Cost of Clean Water-Petroleum Refining," Federal Water Pollution Control Administration, 14-12-100, 1967, U.S. Government Printing Office.) Since it is a typical practice in most refineries to collect all contaminated process wastewater and to combine it into a single wastewater stream and to then treat it in a central treatment facility, it is often unnecessary to deal with the volume and characteristics of each of the component wastewater streams. Also, the nature of the energy-saving process changes considered in this study do not in themselves alter the major wastewater-generating refinery operations. Thus, in assessing the water pollutional implications of alterations to the base case refinery, it is not necessary to examine the individual wastewater sources. Rather, it is necessary to deal with the volume and characteristics of the total refinery wastewater effluent.

b. Wastewater Characteristics

Since petroleum and petroleum products are the major source of pollutants in refinery wastewater, it is not surprising that chemical constituents found in petroleum appear in the wastewater.

Raw refinery wastewater contains large quantities of oil. The oil is present both as free oil (floatable) and as emulsified oil. In addition, water-soluble hydrocarbons, such as phenolic compounds, which are present in the petroleum will also be present in the wastewater.

Crude petroleum contains a variety of sulfur compounds which are removed from the finished product in varying degrees depending on product specifications. Due to the oil/water contacting at various stages of the refining operation, a significant quantity of sulfur compounds enters the wastewater stream. The most objectionable of these sulfur compounds are sulfides, which are typically present in the wastewater as sulfide ions.

Petroleum also contains a number of nitrogenous compounds, and therefore refinery wastewater is typically contaminated with appreciable quantities of ammonia. Small amounts of cyanide compounds are also present.

Carbonaceous and inorganic particulate matter from a variety of sources (e.g., incomplete combustion, soil, and the like) are also present in refinery wastewater, thus contributing to the suspended solids level.

Most of the above-mentioned compounds are oxidizable, and therefore refinery wastewater will exert a chemical oxygen demand (COD).

A certain fraction of the same compounds are biodegradable, and therefore refinery wastewater will also exert a biochemical oxygen demand (BOD_5).

Petroleum also contains a variety of trace heavy metals, e.g., mercury, cadmium, lead, etc., which vary greatly from crude to crude and have not been extensively quantified.

Depending on the type and quantity of corrosion inhibitor used in the non-contact cooling water circuit, the cooling tower blowdown can present pollutional problems of varying significance. The most common types of corrosion inhibitors contain chromate salts. Thus, blowdown from cooling towers can often contain chromium in the form of hexavalent chromium--the more objectionable form.

In the petroleum refining industry, there has been a continuing trend toward wastewater volume reduction. On a gallon per barrel basis, large modern refineries generate far less wastewater than older, smaller refineries. Process improvements, partially motivated by a desire to decrease the volume of wastewater that must be treated, have been largely responsible for this decrease. Improved leak and spill management, coupled with other preventive measures, has also contributed to the overall volume reduction. There has also been a trend toward tighter, non-contact cooling water circuits in addition to more widespread use of air cooling.

Since the free oil fraction of refinery wastewater is readily removable, the first step in any refinery wastewater treatment facility is an oil/water separator, typically referred to as an "API" (American Petroleum Institute) separator. Because of this universal feature, raw wastewater characteristics are almost always measured after the API separator.

Based on a survey of refinery raw waste loads (". . .EPA Development Document - Petroleum Refining . . ." [EPA 440/1-73/014]), a typical unit loading and total waste loadings for the base case refineries are tabulated in Table D-1.

TABLE D-1

TYPICAL BASE CASE RAW REFINERY WASTE LOADS

Wastewater Parameter	Typical Unit Raw Waste Loads	Base Case East and Gulf Coast Raw Waste Load (200,000 BPD)	Base Case West Coast Raw Waste Load (175,000 BPD)
	(lb/Bbl)	(lb/day)	(lb/day)
BOD_5	0.025	5,000	4,375
COD	0.070	14,000	12,250
Suspended solids	0.0096	1,920	1,680
Oil & grease	0.0096	1,920	1,680
Phenols	0.001	200	175
Ammonia (as N)	0.0035	700	612
Sulfide	0.0035	700	612
Flow	33 gal/Bbl	6,600,000 gpd	5,780,000 gpd

Notes: 1) Source: ". . . EPA Development Document - Petroleum Refining . . ." EPA 440/1-73/014.

2) Unit raw waste load is based on the 50% median value for a survey of "low cracking" refineries.

c. Regulatory Constraints

The recommended effluent guidelines for the petroleum refining industry ("Effluent Guidelines and Standards - Petroleum Refining," 40CFR 419 FR, May 9, 1974) place limitations on the following wastewater parameters:

- BOD_5,
- Total suspended solids,
- Chemical oxygen demand,
- Oil and grease,
- Phenolics,
- Ammonia,
- Sulfide,
- Chromium (total and hexavalent), and
- pH.

The guidelines are based on a formula that takes into consideration the refinery process configuration (e.g., "high cracking" vs. "low cracking," etc.) and the refinery size.

Since the basis of comparison used in this study is within the 1985 timeframe, the applicable effluent guideline treatment level is the proposed BATEA (Best Available Technology Economically Achievable) level to be implemented (pending comment and revision) by 1983.

Based on the appropriately adjusted BATEA limitations, treated effluent waste loads have been computed for the base case refineries and are presented in Table D-2.

d. Treatment Technology and Treatment Economics

To achieve the effluent waste loads presented in Table D-2, the following treatment steps are generally considered necessary. These include:

- API oil/water separator,
- Dissolved air flotation,
- Equalization/neutralization,
- Activated sludge system (including aeration basins, clarifiers, and sludge recycle),
- Granular media filtration,
- Carbon adsorption (with thermal regeneration),
- Stormwater collection (from process areas), and
- Chemical feed system (neutralizing chemicals and nutrients for activated sludge system).

TABLE D-2

BASE CASE REFINERY - CHARACTERISTICS OF RAW AND TREATED PROCESS WASTEWATER

Wastewater Parameter	East/Gulf Coast Refinery (200,000 BPD)				West Coast Refinery (175,000 BPD)			
	Raw Wastewater (1)		Treated Effluent (2)		Raw Wastewater (1)		Treated Effluent (2)	
	(lb/day)	(mg/ℓ)	(lb/day)	(mg/ℓ)	(lb/day)	(mg/ℓ)	(lb/day)	(mg/ℓ)
BOD_5	5,000	91	260	4.7	4,375	91	227	4.7
TSS	1,920	35	260	4.7	1,680	35	227	4.7
COD	14,000	254	1,420	25.8	12,250	254	1,243	25.8
Oil & grease	1,920	35	50	0.9	1,680	35	44	0.9
Phenolics	200	3.6	1.0	0.018	175	3.6	0.88	0.018
Ammonia (as N)	700	12.7	316	5.7	612	12.7	277	5.7
Sulfide	700	12.7	4.4	0.08	612	12.7	3.85	0.08
Total chromium	Varies		13(4)	-	Varies		11.4(4)	-
Hexavalent chromium	Varies		0.22(4)	-	Varies		0.19(4)	-
Flow Rate	6,600,000 gpd				5,780,000 gpd			

Notes:
(1) Raw wastewater characteristics derived from EPA Development Document (EPA 440/1-73/014)

(2) Raw wastewater characteristics are for wastewater downstream of the plant is API oil separator.

(3) Treated effluent loadings are based on the Best Available Technology Economically Achievable (BATEA) treatment level for 1983, as defined in "Effluent Guidelines and Standards - Petroleum Refining", 40 CFR 417 FR, May 9, 1974.

(4) Chromium-contaminated cooling water blowdown may be discharged as a separate waste stream.

(5) Waste loads include stormwater runoff from process areas but not from offsite facilities such as tank farms.

To process the sludge produced by the above treatment steps, the following sludge-handling steps can be applied prior to land disposal:

- Sludge thickening,
- Vacuum filtration, and
- Fluid-bed incineration.

While chromium-contaminated cooling tower blowdown is sometimes treated within the central process wastewater treatment facility, separate treatment is becoming more common, because of the effluent limitations placed on chromium. In addition, some of the anticipated energy-saving process changes increase the quantity of cooling water. Thus, by choosing separate treatment for the cooling tower blowdown, these differences will be more clearly highlighted.

The chromium removal system typically consists of the following steps:

- Chromium reduction via sulfur dioxide addition,
- Chromium precipitation (with lime), and
- Landfill of chrome hydroxide sludge.

The chromium reduction takes place in a reaction vessel, while the precipitation is performed in a clarifier. The necessary chemical feed and control equipment are included.

A low-volume, but potentially troublesome, wastewater stream is the purge stream from the Stretford process sulfur recovery system (described under the air pollution control section). This stream contains sodium metavanadate and a diverse collection of organic and sulfur compounds. Its typical composition is shown below:

Constituent	Concentrations (mg/l)
Sodium carbonate	4700
Sodium anthraquinone disulfonate	700
Sodium metavanadate	300
Sodium citrate	300
Sodium thiosulfate	6000
Sodium thiocyanate	6000

While the stream is sometimes mixed with the total process wastewater stream, the presence of vanadium suggests that this is a questionable practice, in terms of potential heavy metals pollution. Since there appears to be no easy way to remove the highly soluble sodium metavanadate, one feasible and relatively foolproof disposal method is chemical fixation, i.e., mixing with a cement-like substance to form a stable solid material. Since the quantity is relatively small, it is reasonable to expect that the cemented material would be disposed of on or near the refinery site.

Wastewater treatment cost estimates, based on the treatment steps previously described have been developed and are presented in Tables D-3 through D-8.

2. AIR POLLUTION

a. Airborne Pollutant Sources and Quantities

The refinery airborne emissions of major consequence are generated by four types of sources:

- Combustion sources - process heaters and boilers;
- Process units - FCC catalyst regenerator and fluid coke;
- Effluent control systems - Claus plant; and
- Storage facilities - gasoline, naphtha, and BTX.

Air emission surveys on refineries have identified the primary pollutants of concern by types of process (Compilation of Air Pollution Emission Factor, Environmental Protection Agency, AP-42, 2nd ed. [1973]). Average emission rates for the principal sources are summarized in Table D-9. Emissions from storage facilities are not included in this tabulation; however, the process changes considered have a negligible effect on storage emissions. Consequently, storage emissions were eliminated from the base line pollutant profiles. Using the emission factor of Table D-9, we determined daily emission rates from major sources within the base line refineries; they are summarized in Tables D-10 through D-12. The sulfur emissions are based upon the actual sulfur balance for each cluster model instead of the representative emission factor. Restrictions were imposed on the refinery fuel oil sulfur level in each base line refinery. These restrictions comply with current regulations for each region represented by the cluster of refineries. The maximum sulfur levels imposed are listed below for each region being studied:

Region	Maximum Sulfur, Wt. %
East Coast	0.6
Gulf Coast	0.9
West Coast	0.7

The sulfur emissions from the Claus unit are based upon 95% recovery of the sulfur in the acid gas. In developing the base line pollution control costs, the control level was increased to 99.5% recovery of sulfur. FCC catalyst regenerator SO_x emissions were not controlled and were not affected by the process changes studied.

b. Pollution Control Costs

The main air pollution control cost assoicated with the base line refineries is for sulfur recovery. Since the refinery fuel sulfur concentration was achieved through gas treating and fuel oil blending to meet existing regulations,

TABLE D-3

PROCESS WASTEWATER TREATMENT COSTS: EAST COAST BASE CASE REFINERY
(Basis: 200,000 BPD, 330 day/yr)

CAPITAL INVESTMENT - $11,600,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
VARIABLE COSTS			
Operating labor (including supervisory & OVHD)	17,520 man-hr/yr	$17.15/man-hr	300,000
Maintenance (including labor & materials)	---	---	464,000
Chemicals			
• Lime	909 tpy	$32.50/ton	29,500
• Phosphoric acid	26 tpy	$440/ton	11,400
• Chlorine	90 tpy	$140/ton	12,600
• Coagulant acid	36,400 lb/yr	$1.00/lb	36,400
• Replacement carbon	123,750 lb/yr	$0.40/lb	49,500
Fuel (gas)	7,425 Btu/yr x 10^6	$1.35/Btu x 10^6	10,000
Electrical power	3,026,000 kWh/yr	$0.030/kWh	90,800
Wastewater treatment sludge disposal			
• Incinerator ash	255 tpy	$5.00/ton	1,300
• Stretford purge water	3,400 tpy	$15.00/ton	51,000
TOTAL VARIABLE COST			$1,056,500
FIXED COSTS			
Depreciation @ 6.25%	---	---	725,000
Taxes & insurance @ 2%	---	---	232,000
TOTAL FIXED COST			$ 957,000
TOTAL ANNUAL COST			$2,013,500
RETURN ON INVESTMENT @ 20%			2,320,000
TOTAL			$4,333,500

Notes:
1. Capital investment adjusted to the 1975 level (ENR Construction Cost Index = 2126)
2. Wastewater treatment costs are based on total implementation (from ground up) of the BATEA level (Best Available Technology Economically Achievable), 1983.
3. Cost estimates are for the specific example, and are in no way intended to represent industry-wide treatment costs.

TABLE D-4

COOLING TOWER BLOWDOWN WASTEWATER TREATMENT COSTS: EAST COAST BASE CASE REFINERY

(Basis: 200,000 BPD, 330 day/yr)

CAPITAL INVESTMENT - $1,382,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
VARIABLE COSTS			
Operating labor (including supervisory & OVHD)	2,200 man-hr/yr	$17.15/man-hr	37,700
Maintenance (including labor & materials)	---	---	55,300
Chemicals			
• Sulfur dioxide	470 tpy	$340/ton	159,800
• Lime	500 tpy	$32.50/ton	16,300
• Sulfuric acid	234 tpy	$51.15/ton	12,000
Fuel (gas)	---	---	---
Electrical power	118,200 kWh/yr	$0.03/kWh	3,500
Wastewater treatment sludge disposal	4,620 tpy (@ 10% solids)	$5.00/ton	23,100
TOTAL VARIABLE COST			$307,700
FIXED COSTS			
Depreciation @ 6.25%			86,400
Taxes & insurance @ 2%			27,600
TOTAL FIXED COST			$114,000
TOTAL ANNUAL COST			$421,700
RETURN ON INVESTMENT @ 20%			$276,400
TOTAL			$698,100

Notes:
1. Capital investment adjusted to the 1975 level (ENR Construction Cost Index = 2126)
2. Wastewater treatment costs are based on a chrome reduction/precipitation system consisting of reaction vessels, clarifiers, chemical feed system and controls.
3. Chromium concentration in untreated cooling tower blowdown is assumed to be 30 mg/l.
4. Cost estimates are for separate treatment of chromium-contaminated cooling tower blowdown. Not all refineries treat cooling tower blowdown separately, and the amount of chromate used varies considerably.

TABLE D-5

PROCESS WASTEWATER TREATMENT COSTS: GULF COAST BASE CASE REFINERY
(Basis: 200,000 BPD, 330 day/yr)

CAPITAL INVESTMENT - $11,600,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
VARIABLE COSTS			
Operating labor (including superv. + OHD)	17,520 man-hr/yr	$17.15/man-hr	300,000
Maintenance (including labor + materials)	-	-	464,000
Chemicals			
• Lime	909 tpy	$32.50/ton	29,500
• Phosphoric acid	26 tpy	$440/ton	11,400
• Chlorine	90 tpy	$140/ton	12,600
• Coagulant acid	36,400 lb/yr	$1.00/lb	36,400
• Replacement carbon	123,750 lb/yr	$0.40/lb	49,500
Fuel (gas)	7,425 Btu/yr x 10^6	$1.27/Btu x 10^6	9,400
Electrical power	3,026,000 kWh/yr	$0.0145/kWh	43,900
Wastewater treatment sludge disposal			
• Incinerator ash	255 tpy	$5.00/ton	1,300
• Stretford purge water	2,086 tpy	$15.00/ton	31,300
TOTAL VARIABLE COST			$989,300
FIXED COSTS			
Depreciation @ 6.25%			725,000
Taxes & insurance @ 2%			232,000
TOTAL FIXED COST			$957,000
TOTAL ANNUAL COST			$1,946,300
RETURN ON INVESTMENT @ 20%			$2,320,000
TOTAL			$4,266,300

Notes:
1. Capital investment adjusted to the 1975 level (ENR Construction Cost Index = 2126)
2. Wastewater treatment costs are based on total implementation (from groundup) of the BATEA level (Best Available Technology Economically Achievable) 1983.
3. Cost estimates are for the specific example, and are in no way intended to represent industry-wide treatment costs.

TABLE D-6

COOLING TOWER BLOWDOWN WASTEWATER TREATMENT COSTS: GULF COAST BASE CASE REFINERY
(Basis: 200,000 BPD, 330 day/yr)

CAPITAL INVESTMENT - $1,306,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
VARIABLE COST			
Operating labor (including superv. + OHD)	2,200 man-hr/yr	$17.15/man-hr	37,700
Maintenance (including labor + materials)			52,200
Chemicals			
• Sulfur dioxide	429 tpy	$340/ton	145,900
• Lime	458 tpy	$32.50/ton	14,900
• Sulfuric acid	215 tpy	$51.15/ton	11,000
Fuel	-	-	-
Electrical power	108,400 kWh/yr	$0.0145/kWh	1,600
Wastewater treatment sludge disposal	4,240 tpy (@ 10% solids)	$5.00/ton	21,200
TOTAL VARIABLE COST			$284,500
FIXED COST			
Depreciation @ 6.25%			81,600
Taxes & insurance @ 2%			26,100
TOTAL FIXED COST			$107,700
TOTAL ANNUAL COST			$392,200
RETURN ON INVESTMENT @ 20%			$261,200
TOTAL			$653,400

Notes:
1. Capital investment adjusted to the 1975 level (ENR Construction Cost Index = 2126)
2. Wastewater treatment costs are based on a chrome reduction/precipitation system consisting of reaction vessels, clarifiers, chemical feed system and controls.
3. Chromium concentration in untreated cooling tower blowdown is assumed to be 30 mg/liter.
4. Cost estimates are for separate treatment of chromium-contaminated cooling tower blowdown. Not all refineries treat cooling tower blowdown separately, and the amount of chromate used varies considerably.

TABLE D-7

PROCESS WASTEWATER TREATMENT COSTS: WEST COAST BASE CASE REFINERY
(Basis: 175,000 BPD, 330 day/yr)

CAPITAL INVESTMENT - $10,636,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
VARIABLE COSTS			
Operating labor (including superv. + OHD)	16,060 man-hr/yr	$17.15/man-hr	275,000
Maintenance (including labor + materials)	-	-	425,000
Chemicals			
• Lime	795 tpy	$32.50/ton	25,800
• Phosphoric acid	22 tpy	$440.00/ton	9,700
• Chlorine	79 tpy	$140.00/ton	11,100
• Coagulant acid	31,800 lb/yr	$1.00/lb	31,800
• Replacement carbon	108,300 lb/yr	$0.40/lb	43,300
Fuel (gas)	6,500 Btu/yr x 10^6	$1.31/Btu x 10^6	8,500
Electrical power	2,648,500 kWh/yr	$0.0211/	55,900
Wastewater treatment sludge disposal			
• Incinerator ash	224 tpy	$5.00/ton	1,100
• Stretford purge water	5,460 tpy	$15.00/ton	81,900
TOTAL VARIABLE COST			$969,100
FIXED COSTS			
Depreciation @ 6.25%			664,800
Taxes & insurance @ 2%			212,700
TOTAL FIXED COST			$877,500
TOTAL ANNUAL COST			$1,846,600
RETURN ON INVESTMENT @ 20%			$2,127,000
TOTAL			$3,973,600

Notes: 1. Capital investment adjusted to the 1975 level (ENR Construction Cost Index = 2126)

2. Wastewater treatment costs are based on total implementation (from ground up) of the BATEA level (Best Available Technology Economically Achievable) 1983

3. Cost estimates are for the specific examples, and are in no way intended to represent industry-wide treatment costs

TABLE D-8

COOLING TOWER BLOWDOWN WASTEWATER TREATMENT COST: WEST COAST BASE CASE REFINERY

(Basis: 175,000 BPD, 330 day/yr)

CAPITAL INVESTMENT - $1,147,000

	Yearly Quantity	Cost Per Unit Quantity	Yearly Cost ($/yr)
VARIABLE COST			
Operating labor (including superv. + OHD)	2,200 man-hr/yr	$17.15/man-hr	37,700
Maintenance (including labor + materials)	-	-	45,900
Chemicals			
• Sulfur dioxide	351 tpy	$340/ton	119,300
• Lime	373 tpy	$32.50/ton	12,100
• Sulfuric acid	176 tpy	$51.15/ton	9,000
Fuel	-	-	-
Electrical power	88,700 kWh/yr	$0.0211/kWh	1,900
Wastewater treatment sludge disposal	3,465 tpy (@10% solids)	$5.00/ton	17,300
TOTAL VARIABLE COST			$243,200
FIXED COST			
Depreciation @ 6.25%			71,700
Taxes & insurance @ 2%			22,900
TOTAL FIXED COST			$94,600
TOTAL ANNUAL COST			$337,800
RETURN ON INVESTMENT @ 20%			$229,400
TOTAL			$567,200

Notes:
1. Capital investment adjusted to the 1975 level (ENR Construction Cost Index = 2126)
2. Wastewater treatment costs are based on a chrome reduction/precipitation system consisting of reaction vessels, clarifiers, chemical feed system and controls.
3. Chromium concentration in estimated cooling tower blowdown is assumed to be 30 mg/liter.
4. Cost estimates are for separate treatment of chromium-contaminated cooling tower blowdown. Not all refineries treat cooling tower blowdown separately, and the amount of chromate used varies considerably.

TABLE D-9

EMISSION FACTORS IN PETROLEUM REFINERIES

Type of Process	Particulates	Sulfur Oxides (SO_2)	Carbon Monoxide	Hydrocarbons	Nitrogen Oxides (NO_x)	Aldehydes	Ammonia
Boilers and Process Heaters							
lb/10^3 Bbl oil burned	840	6,720S	Neg	140	2,900	25	Neg
kg/10^3 liters oil burned	2.4	19.2S	Neg	0.4	8.3	0.071	Neg
lb/10^3 f^3 gas burned	0.02	2s	Neg	0.03	0.23	0.003	Neg
kg/10^3 m^3 gas burned	0.32	32s	Neg	0.48	3.7	0.048	Neg
Fluid Catalytic Cracking Units							
Uncontrolled							
lb/10^3 Bbl fresh feed	242 (93 to 340)	493 (313 to 525)	13,700	220	71.0 (37.1 to 145.0)	19	54
kg/10^3 liters fresh feed	0.695 (0.267 to 0.976)	1.413 (0.898 to 1.505)	39.2	0.630	0.204 (0.107 to 0.416)	0.054	0.155
Electrostatic precipitator and CO boiler							
lb/10^3 Bbl fresh feed	44.7 (12.5 to 61.0)	493 (313 to 525)	Neg	220	71.0 (37.1 to 145.0)	19	54
kg/10^3 liters fresh feed	0.128 (0.036 to 0.175)	1.413 (0.898 to 1.505)	Neg	0.630	0.204 (0.107 to 0.416)	0.054	0.155
Moving-Bed Catalytic Cracking Units							
lb/10^3 Bbl fresh feed	17	60	3,800	87	5	12	6
kg/10^3 liters fresh feed	0.049	0.171	10.8	0.250	0.014	0.034	0.017
Fluid Coking Units							
Uncontrolled							
lb/10^3 Bbl fresh feed	523	NA	Neg	Neg	Neg	Neg	Neg
kg/10^3 liters fresh feed	1.50	NA	Neg	Neg	Neg	Neg	Neg
Electrostatic precipitator							
lb/10^3 Bbl fresh feed	6.85	NA	Neg	Neg	Neg	Neg	Neg
kg/10^3 liters fresh feed	0.0196	NA	Neg	Neg	Neg	Neg	Neg
Compressor Internal Combustion Engines							
lb/10^3 ft^3 gas burned	Neg	2s	Neg	1.2	0.9	0.1	0.2
kg/10^3 m^3 gas burned	Neg	32s	Neg	19.3	14.4	1.61	3.2
Blowdown Systems							
Uncontrolled							
lb/10^3 Bbl refinery capacity	Neg	Neg	Neg	300	Neg	Neg	Neg
kg/10^3 liters refinery capacity	Neg	Neg	Neg	0.860	Neg	Neg	Neg
Vapor Recovery System or Flaring							
lb/10^3 Bbl refinery capacity	Neg	Neg	Neg	5	Neg	Neg	Neg
kg/10^3 liters refinery capacity	Neg	Neg	Neg	0.014	Neg	Neg	Neg
Process Drains							
Uncontrolled							
lb/10^3 Bbl wastewater	Neg	Neg	Neg	210	Neg	Neg	Neg
kg/10^3 liters wastewater	Neg	Neg	Neg	0.600	Neg	Neg	Neg
Vapor recovery or separator covers							
lb/10^3 Bbl wastewater	Neg	Neg	Neg	8	Neg	Neg	Neg
kg/10^3 liters wastewater	Neg	Neg	Neg	0.023	Neg	Neg	Neg

TABLE D-9

EMISSION FACTORS PETROLEUM REFINERIES (Cont.)

Type of Process	Particulates	Sulfur Oxides (SO_2)	Carbon Monoxide	Hydro-carbons	Nitrogen Oxides (NO_x)	Alde-hydes	Ammonia
Vacuum Jets							
Uncontrolled							
lb/10^3 Bbl vacuum distillate	Neg	Neg	Neg	130	Neg	Neg	Neg
kg/10^3 liters vacuum distillate	Neg	Neg	Neg	0.370	Neg	Neg	Neg
Fume burner or waste-heat boiler							
lb/10^3 Bbl vacuum distillate	Neg	Neg	Neg	Neg	Neg	Neg	Neg
kg/10^3 liters vacuum distillate	Neg	Neg	Neg	Neg	Neg	Neg	Neg
Cooling Towers							
lb/10^6 gal cooling water	Neg	Neg	Neg	6	Neg	Neg	Neg
kg/10^6 liters cooling water	Neg	Neg	Neg	0.72	Neg	Neg	Neg
Pipeline Valves and Flanges							
lb/10^3 Bbl refining capacity	Neg	Neg	Neg	28	Neg	Neg	Neg
kg/10^3 liser refining capacity	Neg	Neg	Neg	0.080	Neg	Neg	Neg
Vessel Relief Valves							
lb/10^3 Bbl refining capacity	Neg	Neg	Neg	11	Neg	Neg	Neg
kg/10^3 liter refining capacity	Neg	Neg	Neg	0.031	Neg	Neg	Neg
Pump Seals							
lb/10^3 Bbl refining capacity	Neg	Neg	Neg	17	Neg	Neg	Neg
kg/10^3 liter refining capacity	Neg	Neg	Neg	0.049	Neg	Neg	Neg
Compressor Seals							
lb/10^3 Bbl refining capacity	Neg	Neg	Neg	5	Neg	Neg	Neg
kg/10^3 liter refining capacity	Neg	Neg	Neg	0.014	Neg	Neg	Neg
Miscellaneous (air blowing, sampling, etc.)							
lb/10^3 Bbl refining capacity	Neg	Neg	Neg	10	Neg	Neg	Neg
kg/10^3 liter refining	Neg	Neg	Neg	0.029	Neg	Neg	Neg

Source: "Compilation of Air Pollution Emission Factors," AP-42, Office of Air Quality Planning and Standards, EPA 2nd Edition, April 1973.

TABLE D-10

ANTICIPATED SHIFTS IN REFINERY ENERGY USAGE CAUSED BY NATURAL GAS AND TEL PHASEOUT

	East Coast		Gulf Coast		West Coast	
	1974	1985	1974	1985	1974	1985
Crude Run, 10^3 BPD	188	198	222	218	155	164
Purchased Natural Gas, FOE/day[1]	2,500	---	5,400	---	6,390	---
Refinery Fuel, FOE/day[1]						
Gas	4,810	6,990	5,450	6,880	7,190	7,500
Oil (blended)	7,210	9,640	5,350	10,700	2,140	11,690
Total Fuel, FOE/day[1]	14,520	16,630	16,200	17,580	15,720	19,190
Purchased Steam, 10^6 lb/day	5.6	5.7	0.5	0.5	0.7	0
Electricity, mWh/day	713.7	838.7	744.2	875	766.8	897.3

[1] FOE = Fuel Oil Equivalent, 6.3×10^6 Btu.

TABLE D-11

MAJOR AIRBORNE EMISSIONS FROM BASE LINE REFINERY
Location: East Coast
Year: 1985

Process/Pollutants	Particulates (lb/day)	SO_x (lb/day)	Hydrocarbons (lb/day)	NO_x (lb/day)	Aldehydes (lb/day)	Ammonia (lb/day)
● Combustion (heaters and boilers)	8,979	45,200	2,671	38,085	373	Negligible
- Gas-fired	881	---	1,321	10,129	132	Negligible
- Oil-fired	8,098	---	1,350	27,956	241	Negligible
● Fluid catalytic cracking	2,780	52,800	13,684	4,416	1,182	3,359
● Vacuum distillation	Negligible	Negligible	8,931	Negligible	Negligible	Negligible
● Claus plant	---	2,280	---	---	---	---
Total emissions	11,759	100,280	25,286	42,501	1,555	3,359

TABLE D-12

MAJOR AIRBORNE EMISSIONS FROM BASE LINE REFINERY
Location: Gulf Coast
Year: 1985

Process/Pollutants	Particulates (lb/day)	SO_x (lb/day)	Hydrocarbons (lb/day)	NO_x (lb/day)	Aldehydes (lb/day)	Ammonia (lb/day)
• Combustion (heaters and boilers)	9,855	27,600	2,798	40,999	398	Negligible
- Gas-fired	867	---	1,300	9,969	130	Negligible
- Oil-fired	8,988	---	1,498	31,030	268	Negligible
• Fluid catalytic cracking	3,585	8,400	17,644	5,694	1,524	4,331
• Vacuum distillation	Negligible	Negligible	9,854	Negligible	Negligible	Negligible
• Claus plant	---	680	---	---	---	---
Total emissions	13,440	36,680	30,296	46,693	1,922	4,331

there are no segregated costs associated with the control of sulfur oxides from combustion sources. The true cost is contained in the overall refinery operating costs presented in Chapter IV-C of the report.

To assess the impact of pollution control costs on the implementability of the various process changes, base line air pollution control costs were developed in conjunction with the specific process alternative being assessed. This avoided redundancy in determining base line control costs for sources which were not affected by the process changes.

Since sulfur is the pollutant of greatest consequence and control of this pollutant was required for all options, a range of representative control costs were developed for the two basic control systems--Claus tailgas cleanup and flue gas desulfurization.

(1) Tailgas Sulfur Recovery

In many refinery operations, residue gases containing large quantities of H_2S and SO_2 are generated. A large number of processes have been developed for treating these residue gases and recovering saleable byproducts. For details on these processes, the reader is referred to several of the recent journal publications on this subject. (Cost, Air Regulations Affect Process Choice, B.G. Goar, The Oil & Gas Journal, August 1975.) (Environment Needs Guide Refinery Sulfur Recovery, H.S. Bryant, The Oil & Gas Journal, March 1973.)

Almost all of the refineries in the United States currently operate Claus plants designed to recover sulfur from residue gases. The residue gas is first scrubbed with an amine-type scrubbing system to remove the H_2S. The amine scrubbing solutions are then regenerated to give off a concentrated H_2S stream which is sent to the Claus plant. The Claus plant recovers approximately 95% of the sulfur, the rest typically being flared and vented to the atmosphere. In more recent times, both the Federal EPA and State environmental protection agencies have become concerned with the sulfur emissions from Claus plants and have adopted, or are considering the adoption of, standards for the cleanup of tailgases for Claus plants greater than about 20 tpd of sulfur. This regulation would apply to almost all of the refineries in the United States.

The capital cost for a typical sulfur-recovery technology is shown in Figure D-1. The operating costs for this technology are shown in Figure D-2. These data are based upon the Standards Support and Environmental Impact Document (March 1974) to be published by EPA in the near future. The costs shown in these figures are believed to be representative of the whole class of sulfur recovery processes applicable to refinery tailgases.

Since most refineries already have Claus plants, the costs for recovery of additional sulfur due to increased H_2S generation are the incremental costs of building a slightly larger Claus plant than required under today's refinery strategy. This is the approach we have adopted in considering the new technologies for energy conservation.

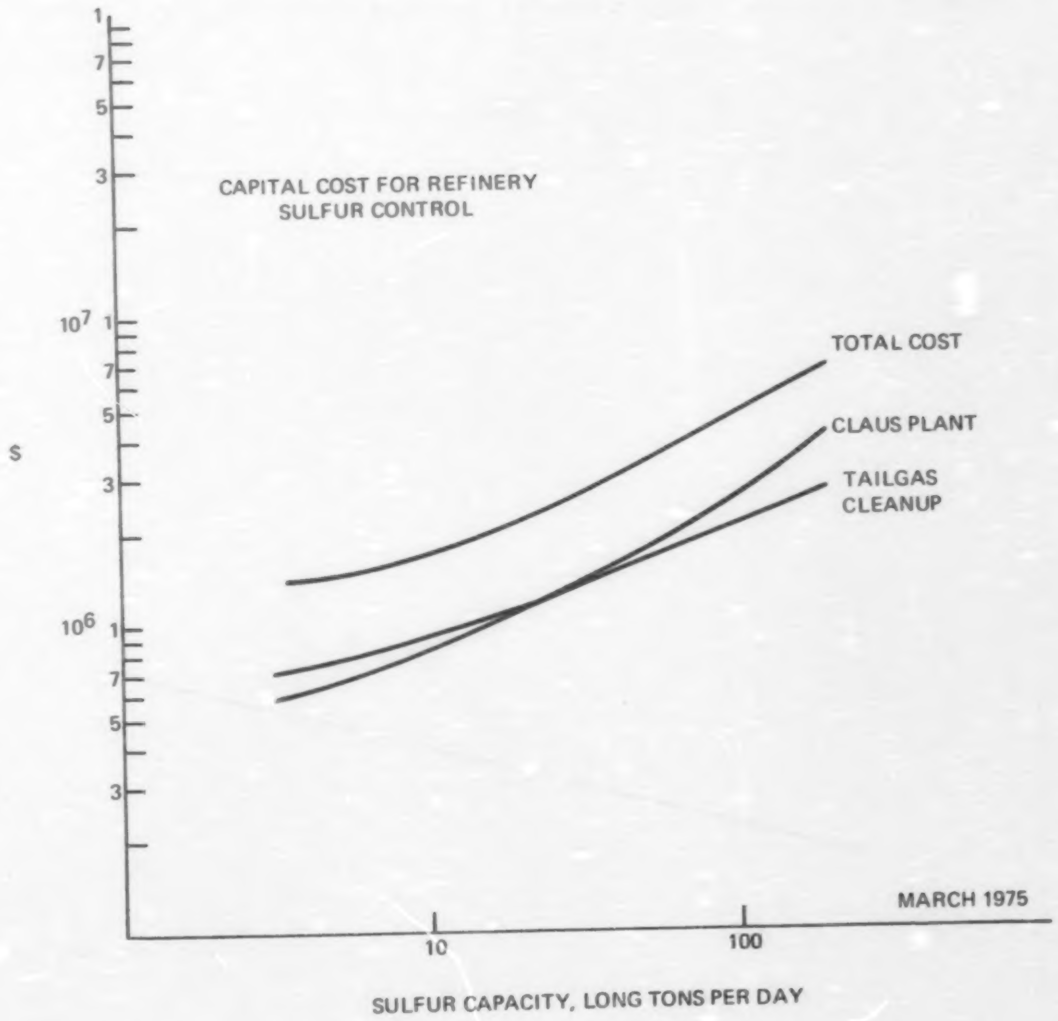


Figure D-1. Capital Cost for a Typical Sulfur - Recovery Technology.

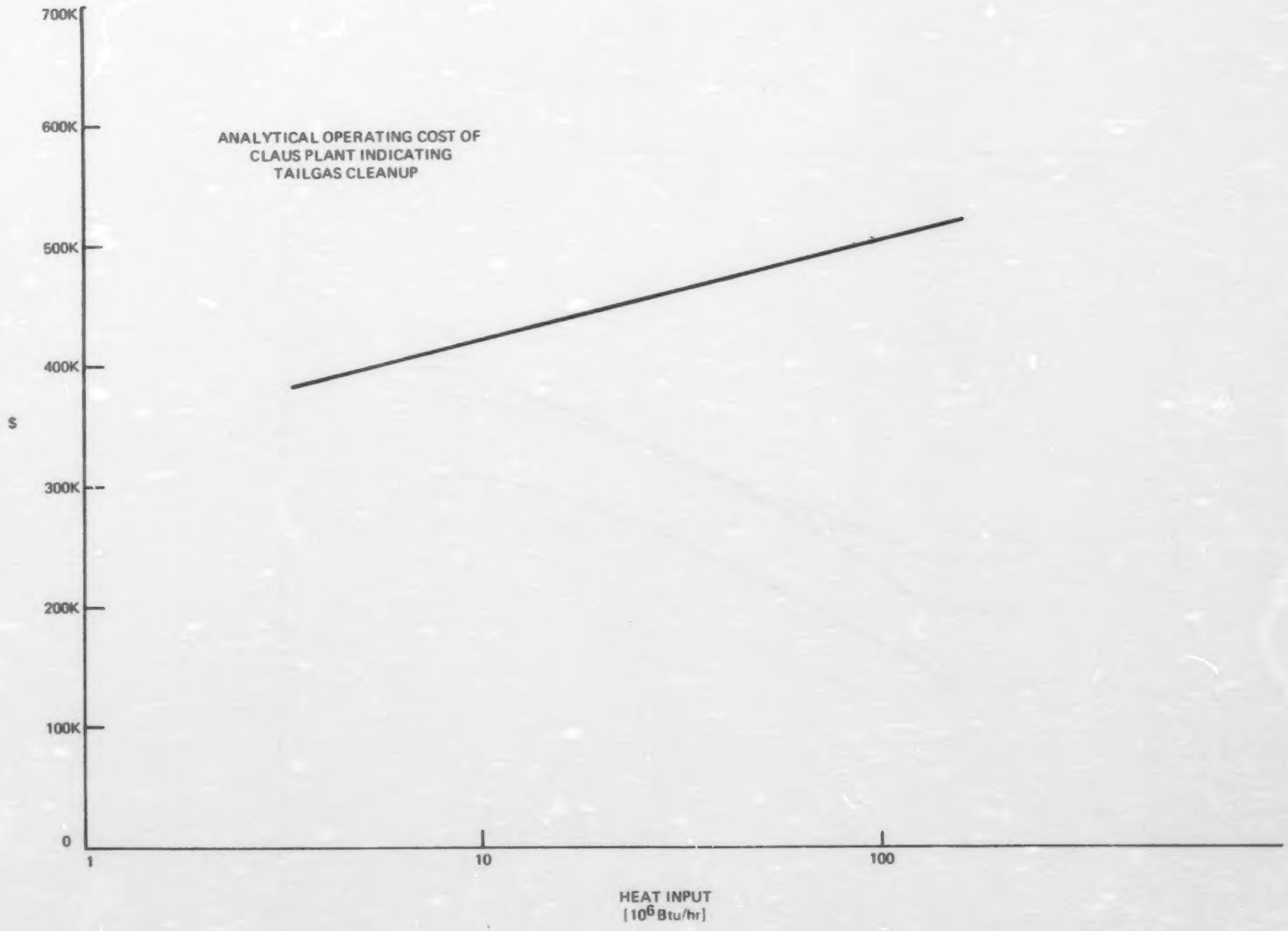


Figure D-2. Operating Costs for Typical Sulfur - Recovery Technology

(2) Flue Gas Desulfurization

One of the greatest environmental problems facing utilities today is the control of SO_2 emissions from boiler stack gases. In recent years several processes have been demonstrated for removing the SO_2 by scrubbing the stack gas with an alkali solution, such as caustic or sodium bisulfite, magnesium hydroxide, lime, and limestone. The more advanced technologies include regeneration of the scrubbing solution by reacting it with limestone or lime to generate calcium sulfite. In other cases, the scrubbing solution is treated to recover sulfur or sulfuric acid using thermal or electrolytic means. The details of these scrubbing systems appear in numerous EPA reports and will not be covered here.

The capital and operating costs for SO_2 scrubbing systems can be calculated from equations reported by Burchard. (Some General Economic Considerations of Flue Gas Scrubbing for Utilities, J.K. Burchard, Control Systems Division, EPA.) For a clear solution (as opposed to lime or limestone) venturi scrubber, the capital cost for controling a new installation using a single scrubber is given by the equation:

$$C_S = 21,600\ Q^{0.65}$$

where (1)

C_S = scrubber cost

Q = heat input 10^6 Btu's/hr

The alkali system necessary for regeneration (based upon a single regeneration system recovering sulfuric acid) is given by the equation:

$$C_R = 3.3 \times 10^6\ [S]^{0.67}$$

where (2)

C_R = regenerations system cost (\$), and

S = sulfur rate (ton/hr).

The total direct cost for this scrubber system would be the sum of costs from Equations (1) and (2). The total system cost would be the direct costs plus indirect costs, estimated at approximately 30% of the direct cost.

Equations (1) and (2) form the basis for calculating the cost for scrubber systems on boilers and process heaters. The capital and operating costs are shown as a function of heat input rate in Figure D-3. A breakdown of the operating costs is shown in Table D-13. Note that refinery process heaters are typically in the range of 50 to 100 million Btu per hour heating rate. Boilers, on the other hand, are usually larger, but rarely are they greater than 500 x 10^6 Btu/hr (less than 50 MW).

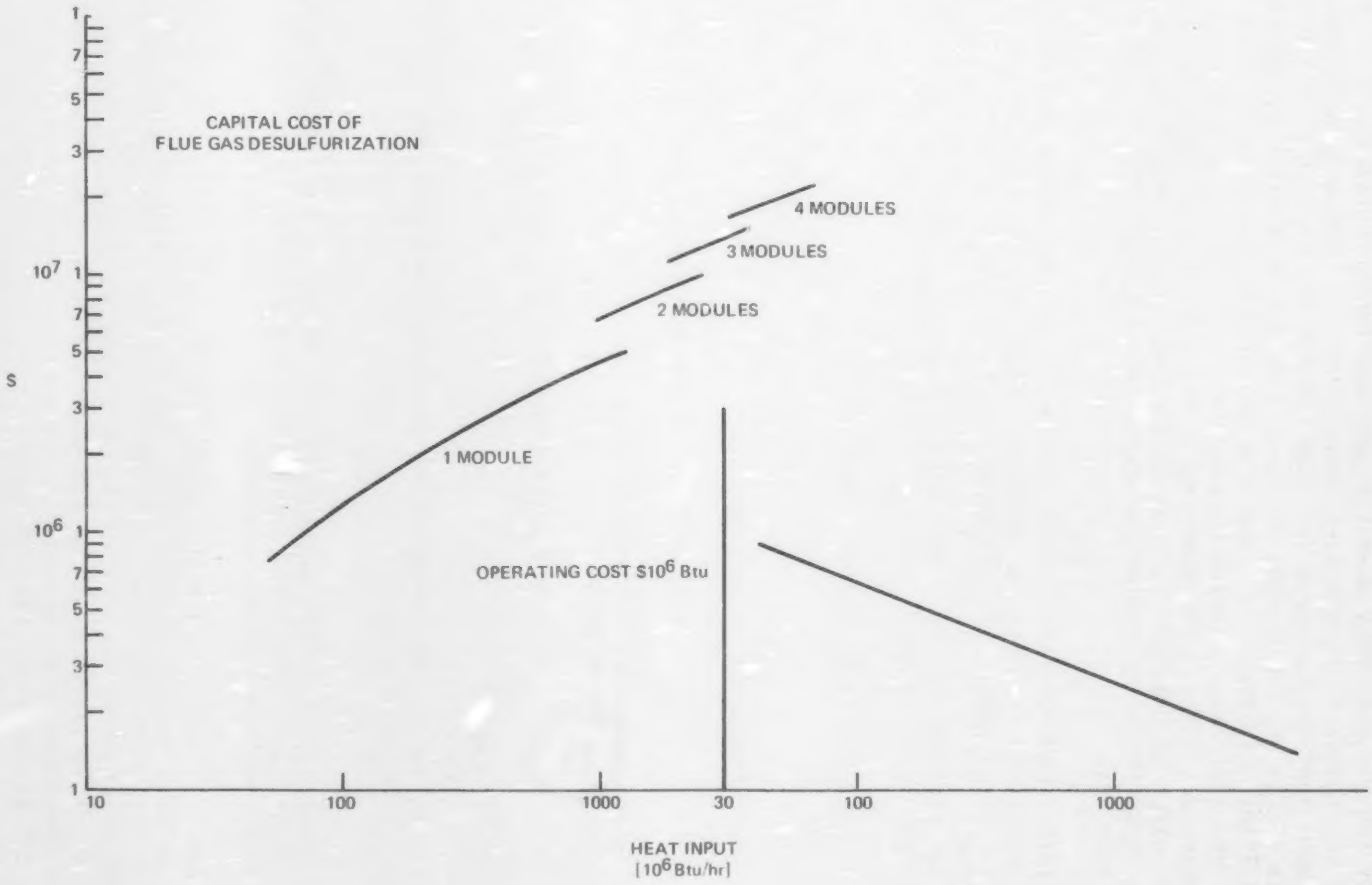


Figure D-3. Capital and Operating Costs vs. Heat Input Rate for Scrubber System

TABLE D-13

AIR POLLUTION CONTROL COST OF LARGE COMBUSTION SOURCES
(Basis: 2% Sulfur)

Heat rate, 10^6 Btu/hr	250	5,250
Modules	1	4
Capital cost, $1000's		
- Scrubber system	782	9,189
- Regeneration system	880	6,770
Total direct cost	1,662	15,959
Indirect costs, @ 30% of direct	499	4,788
Total capital cost, $1000's/yr	2,161	20,747
Operating cost, $1000's/yr		
Variable costs:		
Labor, 1 man/shift (incl. supv. & overhead @ $14.85/hr)	130	130
Maintenance, @ 5% of direct capital	83	798
Utilities		
- Electric power, 240 kWh/10^6 scf @ $0.03/kWh	189	1,656
- Water, 1 gal/10^3 scf @ $0.35/$10^3$ gal	9	80
- Fuel, 2.5 x 10^6 Btu/ton of S @ $2.17/$10^6$ Btu	7	139
Total variable costs	478	2,803
Fixed costs:		
Depreciation, 16 years	135	1,297
Insurance & taxes @ 2% of capital	43	415
Total fixed costs	178	1,712
Total production cost	656	4,515
Return on investment @ 20% of capital	432	4,149
TOTAL ANNUAL COST, $1000's/yr	1,088	8,664
Unit Cost, $/$10^6$ Btu	0.55	0.20

3. SOLID WASTE

A petroleum refinery generates a wide variety of solid waste streams, many of which contain materials on the EPA toxic substances list.

While the total combined treated wastewater effluent from modern refineries is of "relatively" uniform composition and has been extensively quantified, the nature and quantity of solid wastes emanating from refineries are highly variable and still the subject of investigation.

Basically, refinery solid waste streams fall into two main groups: those that are intermittently generated and those that are continuously generated.

Intermittent wastes are generally those that result from cleaning within the process areas and off-site facilities of the refinery. The following are typical intermittent waste streams:

- Process vessel sludges, vessel scale, and other deposits generally removed during plant turnarounds;
- Storage tank sediments; and
- Product treatment wastes, such as spent filter clay and spent catalysts from certain processing units.

The annual volume of refinery intermittent wastes is strongly a function of the individual refinery waste management and housekeeping practices.

Continuous wastes (those requiring disposal at less than 2-week intervals) can be further broken down into two groups: process unit wastes and wastewater treatment wastes.

Major process unit wastes include:

- Coker wastes, such as coke fines from delayed or fluidized cokers, and spilled coke from unloading facilities;
- Spent catalysts and catalyst fines from the fluid catalytic cracking units; and
- Spent and spilled grease and wax wastes from lube oil processing plants.

Wastewater treatment wastes can include:

- Waste biological sludges from activated sludge units; and
- Dissolved air flotation float.

Typically such wastes are dewatered by means of sludge thickeners, coupled with vacuum filters or centrifuges. The dewater sludge can then either be land-disposed or incinerated. Low concentrations of heavy metals are usually present in the sludges, which could affect the level of control required.

A summary description of major refinery solid waste streams is presented in Table D-14.

Using estimates for the total refinery solid waste generation in the these are presented in Table D-15. As can be seen from Table D-15, the base case East and Gulf Coast refineries produce a total solid waste stream that is estimated to be 19,400 tpy. The West Coast refinery produces a total solid waste stream estimated to be 16,700 tpy.

Excluding wastewater treatment sludges (which are included under water pollution control costs), the total yearly solid waste disposal costs, based on a typical unit disposal cost of $5.00/ton, are $91,000/yr for the East and Gulf Coast base case refineries and $80,000/yr for the base case West Coast refinery. It should be stressed that these are very rough estimates and that actual values can vary considerably from plant to plant.

TABLE D-14

PETROLEUM REFINERY SOLID WASTES--SOURCES AND CHARACTERISTICS

Type of Waste	Sources	Description	General Characteristics
Process solids	Crude oil storage, desalter	Basic sediment and water	Iron rust, iron sulfides, clay, sand, water, oil
	Catalytic cracking	Catalyst fines	Inert solids, catalyst particles, carbon
	Coker	Coker fines	Carbon particles, hydrocarbons
	Alkylation	Spent sludges	Calcium fluoride, bauxite, aluminum chloride
	Lube oil treatment	Spent clay sludges, press dumps	Clay, acid sludges, oil
	Drying and sweetening	Copper sweetening residues	Copper compounds, sulfides, hydrocarbons
	Storage tanks	Tank bottoms	Oil, water, solids
	Slop oil treatment	Precoat vacuum filter sludges	Oil, diatomaceous earth, solids
Effluent treatment solids	API separator	Separator sludge	Oil, sand, and any of the above process solids
	Chemical treatment	Flocculant aided precipitates	Aluminum or ferric hydroxides, calcium carbonate
	Air flotation	Scums or froth	Oil, solids, flocculants (if used)
	Biological treatment	Waste sludges	Water, biological solids, inerts
General waste	Water treatment plant	Water treatment sludges	Calcium carbonate, alumina ferric oxide, silica
	Maintenance	Heat exchanger bundle cleaning	Iron rust, sediment, oil

---

Source: "Assessment of Industrial Hazardous Waste Practices, Petroleum Refining--Draft Report," U.S. Environmental Protection Agency, January 1975.

TABLE D-15

BASE CASE REFINERY - CHARACTERISTICS OF RAW AND TREATED PROCESS WASTEWATER

Wastewater Parameter	East/Gulf Coast Refinery (200,000 BPD)				West Coast Refinery (175,000 BPD)			
	Raw Wastewater (1)		Treated Effluent (2)		Raw Wastewater (1)		Treated Effluent (2)	
	(lb/day	(mg/l)	(lb/day	(mg/l)	(lb/day	(mg/l)	(lb/day	(mg/l)
BOD_5	5,000	91	260	4.7	4,375	91	227	4.7
TSS	1,920	35	260	4.7	1,680	35	227	4.7
COD	14,000	254	1,420	25.8	12,250	254	1,243	25.8
Oil & grease	1,920	35	50	0.9	1,680	35	44	0.9
Phenolics	200	3.6	1.0	0.018	175	3.6	0.88	0.018
Ammonia (as N)	700	12.7	316	5.7	612	12.7	277	5.7
Sulfide	700	12.7	4.4	0.08	612	12.7	3.85	0.08
Total chromium	Varies		13(4)	-	Varies		11.4(4)	-
Hexavalent chromium	Varies		0.22(4)	-	Varies		0.19(4)	-
Flow Rate	6,600,000 gpd				5,780,000 gpd			

Notes:
(1) Raw wastewater characteristics derived from EPA Development Document (EPA 440/1-73/014)

(2) Raw wastewater characteristics are for wastewater downstream of the plant API oil separator.

(3) Treated effluent loadings are based on the Best Available Technology Economically Achievable (BATEA) treatment level for 1983, as defined in "Effluent Guidelines and Standards - Petroleum Refining", 40 CFR 417 FR May 9, 1974.

(4) Chromium - contaminated cooling water blowdown may be discharged as a separate waste stream.

(5) Waste loads include stormwater runoff from process areas but not from offsite facilities such as tank farms.

APPENDIX E

ROSTER OF PROCESS OPTIONS OR CHANGES IN INDUSTRIAL PRACTICE

The main body of the report describes in detail the options investigated, the criteria for their implementation, and the base line technology for comparison.

As an initial point in the refinery study ADL interviewed major oil companies and process licensors to discuss with them their views concerning process changes which will be made in existing refineries during the 1975 to 1985 period. Using information obtained from these interviews plus other sources of technical information available to ADL, we selected the process options described in the main body of the report as those most likely to be implemented into existing refineries.

A complete list of the major process options considered is given below.

1. Upgrading of asphaltic materials (heavy residium):
    a. via partial oxidation to make hydrogen,
    b. via Flexicoking, and
    c. via residual desulfurizing;
2. Complete internal power generation, i.e., no purchased power;
3. New sources of hydrogen manufacture from:
    a. residual material via partial oxidation (same as 1(a) above),
    b. coal, and
    c. naphtha;
4. Use of a high-sulfur fuel, plus flue gas desulfurization;
5. Exporting methane, ethane, and larger quantities of LPG from the refinery, coupled with increased internal consumption of heavier material for refinery fuel;
6. Conversion of LPG to gasoline;

7. Upgrading of hydrogen purity via:

   a. installation of the DuPont membrane separation process, and

   b. installation of a computer-controlled hydrogen makeup bleed system;

8. Conversion of some sulfuric acid alkylation plants to HF alkylation plants;

9. Blending of methanol into the gasoline pool;

10. Larger market penetration of diesel automobiles and, hence, larger demand for diesel fuel; and

11. Use of coal as refinery fuel for internal power generation.

The first four alternatives were finally selected for detailed evaluation, excepting alternatives 3b and 3c. The remaining options are briefly discussed below.

5. Exporting Methane, Ethane, and LPG

One possible scenario discussed during the interviews was the question: "Would decontrol of natural gas pricing or end-use restrictions ultimately result in the sale of refinery gas to utilities?" It is conceivable that the price of natural gas could rise to such a level as to appear to make it attractive for a refinery to export natural gas. In general, we would not expect this to happen. Natural gas and/or light fuel is essential to the control of many process furnaces, e.g., the catalytic reformer furnaces. If all the fuel gas were exported, it would have to be replaced (presumably) with fuel oil. This would present the refiner with difficult temperature control problems in a good many furnaces.

An additional comment made to ADL during the interviews was that, if export of refinery gas becomes economic, then SNG will become economic at the same time and would be the preferred route. A further difficulty with export of refinery gas for pipeline use is its variability in heating value--from 550 to 1100 Btu over relatively short time spans. However, over the fence arrangements to supply gas to gas-dependent industries would be feasible.

6. Conversion of LPG to gasoline

This process has been suggested several times in the past as a method for gasoline production, and is under active consideration by some companies, particularly in Australia. The economics do not appear attractive in the United States at this time.

7. Upgrading of Hydrogen Purity

Considerable quantities of hydrogen are manufactured within the refinery, particularly in the catalytic reformers. This hydrogen is typically low purity, relative to purity of manufactured hydrogen. Hydrogen purity in the hydrotreating hydrodesulfurizing reactors is one of the key variables affecting throughput, conversions, etc. Upgrading of the low-purity hydrogen produced by the catalytic reformers would potentially improve the performance of an existing hydrotreater/hydrodesulfurizer. For example, heavier stocks could be fed to an existing unit, the conversion could be altered, etc. Three methods have been studied to upgrade hydrogen purity:

a. Cryogenic separation,

b. DuPont's permasep process (membrane), and

c. Palladium diffusion.

The palladium diffusion process is not likely to be applied in a refinery, because the process operation is sensitive to sulfur in the feed. Most companies interviewed expected a demonstration unit of the DuPont process to be installed within the next year or so. The choice between this process (assuming satisfactory operation) and cryogenic separation is one of economics. The membrane separation process could have a significant energy advantage.

To date, hydrogen purity upgrading is not widely practiced.

The control of the hydrogen bleed rate (to prevent the buildup of inerts) and the hydrogen production rate and purity are critical to minimizing costs. As the price of hydrogen increases, more companies will likely opt for computer control of this system. At least one company now has such a system installed. The net result is not any direct process change, but can represent a significant savings in costs and energy.

8. Conversion of Existing Sulfuric Acid Alkylation Plants to HF Alkylation

Alkylation is an important process for the manufacture of high octane gasoline blending components. Alkylation is the reaction between olefins (C3, C4, C5) produced from the FCC unit and isobutane to produce alkylate. A catalyst, either sulfuric acid or HF acid, is required. If sulfuric acid is used, the reactor temperature is below ambient; hence, refrigeration is required. HF alkylation does not require refrigeration. Because refrigeration is energy-intensive, there might be a potential for energy savings by switching from sulfuric to HF alkylation.

The companies interviewed did not think this switch would be likely to occur, particularly because HF is a very dangerous chemical and plant operation/safety is more difficult.

9. Blending of Methanol into the Gasoline Pool

Methanol is a high-octane material which could potentially be used as an automotive fuel, either in the gasoline pool as a blending component, or as a separate (pure) fuel.

Some of the properties of methanol and a typical gasoline are shown below for comparative purposes (The Oil & Gas Journal, December 17, 1973, pg 71).

Property	Methanol	Gasoline
Energy Content (gross) Btu/gal	64,700	120,000
Octane number	100+	100
Stoichiometric air/fuel ratio (lb/lb)	6.5	15.05
Vapor Pressure, 60°F, psi	4.6	3-4
Heat of Vaporization, Btu/lb	473	130
Density, lb/gal at 60°F	6.64	6.2
Estimated mileage % of gasoline	55-60	100

The same article presented estimated emission performance results for a Gremlin automobile, showing methanol and gasoline to be roughly comparable.

The use of methanol as a blending component in gasoline has been extensively studied by the companies interviewed. They were almost unanimous in the view that methanol is not an economic choice. If water is present anywhere in the distribution system, even in part per million quantities, phase separation can easily occur. The practical difficulties in guaranteeing the almost total absence of water are, in the opinion of the companies interviewed, too great to allow use of methanol. Even humid air entering the tank during filling/emptying is too much water.

Also because of the vapor pressure characteristics of methanol in a typical gasoline pool, butanes/pentanes must be removed. These latter compounds have a higher energy content per gallon than methanol and are less costly.

Methanol can be very corrosive to parts of the fuel system in the automobiles--the carburetor and plastic parts in particular. In addition, the carburetors would have to be modified for methanol operation.

Another objection raised is that formaldehyde is one of the combustion products of methanol.

Several of the companies did point out that pure methanol can be used in fleet service--taxis, etc., but questioned whether OSHA standards could be met. Methanol can also be burned in turbines.

However, all the companies interviewed stated that their studies indicate the cost of methanol to be too high (except for production in the Mid-East from excess methane) to be competitive. Feedstocks investigated included coal, waste, and cellulose conversion. Most studies reported that it is cheaper to ship LNG rather than methanol at distances greater than 5000 miles.

10. Increased Market Penetration of Diesel Autos

The Department of Transportation is currently funding a study in this area, the results of which should be published in 1976. The oil companies do not believe that diesel-powered automobiles will be a significant market factor before 1985.

11. Use of Coal as a Refinery Fuel

Except for possible use in a utility boiler within the refinery, coal is not likely to be used as a fuel source in refineries. The oil companies do not believe the technology is available to ensure adequate temperature control of a process furnace when coal is used as the heat source. However, many companies are continuing to study this problem.

APPENDIX F

GLOSSARY

Alkylation. A refinery process which reacts light C3/C4 olefins with isobutane (both of which are in the gas phase at ambient conditions) to produce high octane gasoline blending components. This process uses either sulfuric acid or hydrofluoric acid as a catalyst.

Amine Sulfur Removal. DEA or MEA scrubbing for removal of acid gases from refinery fuel gas prior to combustion.

Atmospheric bottoms. The first stage in refinery processing is normally distillation of the crude oil at atmospheric pressure into its natural boiling point components. The bottoms product from this distillation is referred to as "atmospheric bottoms" and normally has an initial boiling point of approximately 650°F. This product is usually suitable for residual fuel oil blending or can be a feedstock to subsequent processes for conversion into lighter fractions.

Aromatics. A class of hydrocarbons whose structure contains at least one unsaturated ring compound containing six carbon atoms. Benzene is the simplest hydrocarbon within this category, and toluene/xylenes are other common aromatics. This class of hydrocarbons exhibits very high gasoline octane numbers.

Asphalt. Bottoms from vacuum flashing, 975-1050°F IBP.

Beavon/Stretford Process. A sulfur removal process generally applied to Claus plant tailgas streams. The process involves catalytic conversion of SO_x to H_2S prior to reduction to elemental sulfur.

Bunker fuel. A grade of residual fuel oil used to power ships.

Catalytic cracking. A refinery process widely practiced in the United States in which heavy petroleum fractions are converted into light components by contact with a catalyst at high temperatures (nearly 1,000°F). This process primarily produces high-octane gasoline blending components. It is a cheaper process than "hydrocracking" defined below, but produces a less attractive yield of products.

Claus Process. A sulfur recovery process which employs thermal and catalytic conversion of SO_x and H_2S to elemental sulfur; widely employed for refining sulfur recovery.

Coking. A refinery process which converts heavy residual petroleum fractions (such as the bottoms product from vacuum distillation) into lighter products and petroleum coke. The petroleum coke yield typically represents about 25% of the feed.

Crude unit overhead stream. The lightest boiling components in crude oil which are recovered in the overhead product by the atmospheric distillation of crude oil. Normally, this stream will contain all the very light boiling hydrocarbons, up through the gasoline boiling range. The distillation end point for this stream can vary between 200° - 400°F.

Deep vacuum flashing. Often the second stage in refinery processing (particularly in the United States) is distillation of the atmospheric bottoms defined above under low pressure (vacuum) conditions. The term "deep vacuum flashing" refers to this process at severe conditions of high temperature/ low pressure operation to produce the maximum yield of distillate products from this process.

Distillate fuel. A class of petroleum products with boiling ranges between approximately 350°F and 700°F which have been produced as an overhead (or distillate) stream in a refinery distillation process. The major products within this category are: aviation turbine fuel, kerosene, No. 2 fuel oil, and diesel fuel.

Flexi-coking.®*A refinery process which is more severe than "coking." In this process, the petroleum coke is "gasified," by a reaction with steam and air to reduce the net coke production to a very small percentage of the original feedstock.

H-Oil.® A proprietory residual hydrocracking process, licensed by Cities Service and HRI, which uses an ebullating catalyst bed. The advantage of this process is that it allows catalyst additions and withdrawals to be made to the system during operation and thus allows processing of poorer quality (such as high metal) feedstocks without excessive catalyst deterioration.

Hydrocracking. A refinery process which reacts heavy liquid petroleum fractions in the presence of a catalyst, in a hydrogen-rich environment. This is a more severe processing operation than "hydrotreating" in that high yields of lighter boiling products are obtained in addition to sulfur removal. The process is characterized by high operating pressures (2,000 psig), high hydrogen consumption, and high capital costs, but it produces very good yields of high quality products.

Hydroskimming. A class of refineries which contain no conversion processes to convert intermediate streams to different boiling fractions Crude oil is separated by distillation into its "natural" yield boiling components. The refinery can include extensive processing to improve product qualities.

Hydrotreated. (Hydro desulfurization) A refinery process in which a hydrocarbon stream is "treated" by chemical reaction caused by contact with a catalyst in a hydrogen atmosphere. The main purpose of this process is to reduce the sulfur content of the hydrocarbon stream, but the process also removes nitrogen and improves other product qualities.

Refinery Gas. A mixture of gases, including carbon monoxide, hydrogen, methane and ethane, produced as byproducts in the conversion of crude petroleum.

TABLE F-1

WEIGHT CONVERSIONS

Unit	Equivalent value
One short ton	2,000 pounds
One metric ton	2,204.6 pounds
One long ton	2,240.0 pounds
One cubic centimeter lead (TEL)	1.06 grams

TABLE F-2

VOLUME CONVERSIONS

Unit	Equivalent value
One imperial gallon	1.201 U.S. gallons
One liter	0.264 U.S. gallons
One U.S. barrel	42.000 U.S. gallons
One cubic meter	264.173 U.S. gallons
One cubic foot	7.481 U.S. gallons

TABLE F-3

NOMENCLATURE

B/SD Bbls/SD	Barrels per stream day
BTU	British Thermal Unit
FGD	Flue gas desulfurization
FOE	Fuel oil equivalent
IBP	Initial Boiling Point
LV	Liquid volume
10^3 B 10^3 bb/s	Thousands of barrels
10^3 B/CD	Thousands of barrels per calendar day
10^3 B/SD	Thousands of barrels per stream day
10^3 BPY	Thousands of barrels per year
10^3 KWH	Thousands of kilowatt hours
10^3 KWH/CD	Thousands of kilowatt hours per calendar day
10^3 lb	Thousands of pounds
10^6 B	Millions of barrels
10^6 B/CD	Millions of barrels per calendar day
10^6 BPY	Millions of barrels per year
10^3 SCF	Thousands of standard cubic feet
10^6 SCF	Millions of standard cubic feet
PAD	Petroleum Administration for Defense (Districts)
POX	Partial Oxidation
PPM	Parts per million
SCF	Standard cubic feet
TEL	Tetra Ethyl lead
\$/B/SD	Dollars per barrel per stream day
$\$10^6$	Millions of dollars
FCC	Fluid catalytic cracker
HT	High temperature
LT	Low temperature
HDU	Hydrodesulfurization unit
TEL	Tetraethyl lead
OCS	Outer continental shelf